PB-238 291

RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS

T. G. Shea, et ál

Bacardi Corporation

Prepared for:

Environmental Protection Agency

July 1974

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EPA-660/2-74-074
July 1974

RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS

By

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Project 800935
Program Element 1BB037
Roap/Task 21 BAC 13

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OFFICE OF RESEARCH AND DEVELOPMENT
U. S. ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D. C. 20460

For sale by the Superintendent of Documents, U.S. Government Printing Office, Washington, D.C. 20402

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ABSTRACT

The rum distillery slops stream, i.e., the underflow produced in the distillation of the fermented molasses mixture, typically contains 70 to 100 gm/l COD, 75 to 85 gm/l dissolved solids, and 3 to 10 gm/l suspended solids, and typically contributes two-thirds of the wastewater flows and over 95 percent of the organic emissions, in the wastewater streams emanating from rum distillery operations.

The general objectives of the present study were to develop an anaerobic digestion process for the treatment of the slops stream at the pilot scale, and to establish design criteria for the full-scale application of the process. Both bench and pilot-scale experimental studies were conducted with the anaerobic contact process flow sheet (incorporating biomass recycle) to permit determination of the Monod kinetic constants and the kinetic relationships describing the anaerobic treatment of the slops. The process kinetics were used to examine the operating and performance characteristics of a plant-scale application (predicted effluent quality, mixed liquor volatile suspended solids concentrations, methane production rates, etc.). A process flow sheet was established and design criteria developed as the basis for estimating the cost of a plant-scale installation. The process kinetic relationships and the economic analysis were used to structure a cost-performance relationship to examine tradeoffs between cost, performance, and selected design variables.

The results of the above indicate the feasibility of the anaerobic contact process for treatment of rum distillery slops as follows:

1. Capability to produce an effluent containing less than 30 gm/l of COD at solids retention times greater than 40 days, in the treatment of a slops stream containing from 70 to 100 gm/l of COD.
2. Range of total annual costs (including amortized, operating, and maintenance costs) varying from \$3.74 per cu m treated at a design capacity of 190 cu m/day (\$14.18 per 1,000 gallons at 50,000 gpd) to \$2.13 per cu m treated at a design capacity of 1,140 cu m/day (\$8.07 per 1,000 gallons at 300,000 gpd).

SELECTED WATER RESOURCES ABSTRACTS
INPUT TRANSACTION FORM

1. Report No.
2.
3. Accession No. PB 238 291

4. Title
Rum Distillery Slops Treatment By Anaerobic Contact Process

5. Report Date
6.
8. Performing Organization Report No.

7. Author(s)
Shea, T. G., E. Ramos, J. Rodriguez, and G. H. Dorion

10. Project No.
800935

9. Organization
Bacardi Corporation, San Juan, Puerto Rico

11. Contract/Grant No.

13. Type of Report and Period Covered

12. Sponsoring Organization **Environmental Protection Agency**

15. Supplementary Notes
Environmental Protection Agency report number, EPA-660/2-74-074, July 1974

16. Abstract

An investigation was made of the feasibility of rum distillery slops treatment by the anaerobic contact process. Both bench and pilot-scale experimental studies were conducted to permit determination of the Monod kinetic constants for the anaerobic treatment of neutralized, phosphorus-amended rum slops. The settling characteristics of mixed liquor suspended solids from the anaerobic contact unit were determined using a moderately cationic, high molecular weight polymer as a coagulant aid. The experimental results, developed with the bench and pilot units, were used to define a process flow sheet, detailed design criteria, and an economic analysis for full-scale applications of rum distillery slops treatment, at design capacities varying from 190 cu m/day (50,000 gal/day) to 1,140 cu m/day (300,000 gal/day).

17a. Descriptors
*Industrial wastes, *Anaerobic digestion, *Energy conversion, Fermentation, Methane

17b. Identifiers
*Rum slops, Monod kinetics, anaerobic treatment

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INFORMATION SERVICE
US Department of Commerce
Springfield, VA. 22151

17c. COWRR Field & Group 05D

18. Availability
19. Security Class. (Report)
20. Security Class. (Page)
21. No. of Pages 109
22. Price MF $2.25

Send To:
WATER RESOURCES SCIENTIFIC INFORMATION CENTER
U.S. DEPARTMENT OF THE INTERIOR
WASHINGTON, D. C. 20240

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Institution W. E. Gates and Associates, Inc.

3. The recovery of methane as an energy byproduct of anaerobic contact treatment of rum distillery slops can, at current energy costs, reduce the above unit treatment costs from one-third at the 190 cu m/day (50,000 gpd) capacity to two-thirds at the 1,140 cu m/day (300,000 gpd) capacity.

Additional treatment of the effluents of the anaerobic contact process will be required to attain effluent quality levels currently defined as BPTCTCA (best practicable control technology currently available) by the Environmental Protection Agency.

This report was submitted in fulfillment of Project 800935 by Bacardi Corporation under the partial sponsorship of the Environmental Protection Agency. Work was completed as of December 1973.

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ACKNOWLEDGMENTS

Of the many individuals in Bacardi Corporation who contributed to the project effort, special acknowledgments are extended to: Mr. William E. Paterson, Vice President and General Manager, for his extensive support in all aspects of the project. Mr. Evelio Gonzalez assisted in the design and set-up of the pilot plant, and Mrs. Nydia Cancel and Mr. Jorge Bracero of Bacardi Corporation provided valuable assistance in the laboratory operations. Dr. John F. Andrews of Clemson University, Dr. A. D. Carr of the University of Capetown, South Africa, and Dr. Paul Smith of the University of Florida, served as Special Consultants to the project.

The firm of W. E. Gates and Associates, Inc. served as Project Consultants, in the latter stages of the effort, and the contributions of several of its staff members are acknowledged: Mr. John D. Stockton and Mr. T. J. Smith, who designed and conducted the pilot plant solids separation tests; Mr. J. E. Hensley, who assisted in data analysis and presentation; and Mrs. Bitsy Smith, for editing and typing the report.

The contributions of Mr. Eric Jerome of Engineering-Science, Inc., with whom the senior author was associated during the early stages of the project, are acknowledged.

The participation and support of Mr. H. George Keeler and Mr. K. A. Dostal of EPA in all phases of the project effort are gratefully acknowledged.

SECTION I

CONCLUSIONS

The conclusions of the present investigation of rum distillery slops treatment by the anaerobic contact process are as follows:

1. Rum distillery slops are amenable to treatment by the anaerobic contact process; effluents containing less than 30 gm/l of COD are produced at solids retention times greater than 40 days in the treatment of slops streams containing from 70 to 100 gm/l of COD.
2. The Monod model describes the response of the anaerobic contact process for the treatment of rum distillery slops even when engineering-type parameters such as VSS (volatile suspended solids) and COD (chemical oxygen demand) are used as measures of viable biomass and limiting substrate, respectively, and is a useful tool in engineering design and analysis of full-scale applications
3. The operational relationships between effluent quality and the primary variables θ_T (hydraulic residence time) and C^o (recycle factor) provide a unifying basis for examining the design problem and establishing operating requirements in the application of the anaerobic contact process for treatment of rum distillery slops
4. The recovery of methane as an energy byproduct of rum distillery slops treatment by the anaerobic treatment process can reduce the unit treatment costs ($/unit volume treated) in a plant-scale installation, at current energy costs, by at least one-third at a design capacity of 190 cu m/day (50,000 gpd) and as much as two thirds at a design capacity of 1,140 cu m/day (300,000 gpd).
5. Additional treatment of effluents from the anaerobic contact process (for removal of COD, suspended solids, and color) will be required in order to attain the effluent quality levels currently defined as best practicable control technology currently available by the Environmental Protection Agency.

SECTION II

RECOMMENDATIONS

Based on the results of the present investigation as documented herein, it is recommended that:

1. The anaerobic contact process, incorporating the recovery of methane as an energy byproduct, be considered as a viable alternative for the treatment of rum distillery slops
2. The development of criteria for a full-scale application of the anaerobic contact process for rum distillery slops treatment be based on (a) the transferable kinetic characterization of the response of the anaerobic contact process, and (b) the framework for development of design criteria to meet specific effluent quality objectives, as presented herein

Specific design criteria are recommended herein (Section VII) for the example design of an anaerobic contact process application to achieve, as a treatment objective, an effluent quality of 30 gm/l COD, equivalent (for a raw slops COD concentration of 100 gm/l) to a soluble COD removal efficiency of 70 percent. The recommended design criteria for the key elements of the process train, as based on an evaluation of experimental results presented in Section VI, are as follows:

1. θ_T/C^o(sludge age) @ 40 days
2. θ_T (hydraulic residence time) @ 15 days, and C^o (recycle factor) @ 0.375
3. Solids loading rate to clarifier @ 100 kg/day/sq m (20.5 lbs/day/sq ft) on overflow basis
4. Lime addition (standby basis) @ 4 kg CaO per cu m/day of slops flow (33.3 lb @ CaO/1,000 gpd)
5. Phosphorus addition (to balance phosphorus requirements for anaerobic biological treatment) @ 0.3 kg P per cu m/day slops flow (2.5 lb P/1,000 gpd)

6. Polymer addition (for conditioning of mixed liquor solids prior to separation) @ 15 gm per cu m/day of slops flow (0.125 lb/1,000 gpd)

In conjunction with process application at the full-scale, additional research is recommended in the following areas:

1. Evaluation of alternative unit process technologies for solids separation (e.g., centrifugation and flotation) and sludge processing, such that cost-performance tradeoffs can be examined in the selection of a separation system.
2. Evaluation of the nutritional value of byproduct sludges as a feed extender.
3. Evaluation of the effectiveness of in-stream process control measures, such as sulfide stripping from the recycled gas stream to control mixed liquor sulfide concentrations, or CO_2 stripping from the recycled gas stream to control mixed liquor pH.
4. Evaluation of alternative strategies for minimizing the time delay in re-startup of an anaerobic contact treatment system at the end of seasonal rum production shutdowns; alternative strategies that merit evaluation include:
    a. Minimizing the degree of reduction in raw slops supply by staggering the shutdown of sectors of the rum production facility
    b. Holding in reserve a quantity of slops to be used to re-start feeding in advance of the resumption of full production
    c. Using an alternative feed supply, such as raw municipal primary sludge, during the shutdown/slowdown period
5. Evaluation of the efficacy of the anaerobic treatment of raw municipal sludges jointly with rum distillery slops, allowing the beneficial use of the nutrient release and alkalinity production associated with the digestion of raw municipal sludges in the treatment of the rum distillery slops.

SECTION III

INTRODUCTION

STUDY OBJECTIVES

The following objectives were established for the study:

1. General objectives
    - a. to develop an anaerobic digestion process for the treatment of the slops stream on a pilot scale
    - b. to establish design criteria for the full-scale application of the process
2. Specific objectives
    - a. to refurbish, install, and operate an existing pilot plant for the treatment of slops
    - b. to determine the process response characteristics and operating boundaries for the anaerobic digestion process and the variations in process efficiency as a function of operating parameters
    - c. to define at the bench scale an information base necessary to design the above pilot scale experimentation
    - d. to verify the applicability of proposed kinetic relationships for the anaerobic treatment of slops, and to establish values of the kinetic constants for these relationships
    - e. to establish design criteria and to develop a preliminary engineering design for the full-scale process
    - f. to estimate capital, operation, and maintenance costs for a plant scale installation

BACKGROUND

The production of rum is accomplished by age-old processes that, in their modern application, are similar in function throughout the industry. The basic sequence of steps in rum production consist of: the mixing of molasses, water, nutrients, and antifoamant; acidification of the above mixture; propagation of the biomass used in the fermentation;

the fermentation itself; and the distillation of the ferment. The distilled spirits are transferred sequentially to holding tanks, to oak barrel filling, to the aging warehouse, and (after a legally-defined period of aging) to barrel emptying operations and thence to the bottling line, in the typical operation. The above rum production and bottling operations are often conducted on a seasonal basis, with shutdown periods of one to two months each year for equipment maintenance and replacement and other reasons.

The basic types of process waste streams generated in the production of rum include:

1. The slops stream, i.e., the underflow produced in the distillation of the fermented molasses mixture.
2. Barrel washings.
3. Cooling tower and boiler plant blowdown.
4. Regenerant water from water treatment facilities, analytical laboratory wastewaters, and fermenter washdown.

Presented in Table 1 is a profile of (1) the total facility wastewater generation per proof gallon produced from all of the above process streams, and (2) an analysis of the percent contribution to the total wastewater generation by each type of process stream. In review of these data:

1. The slops discharge constitutes 66 percent of the waste flow, over 98 percent of the BOD and COD emissions, over 90 percent of the solids emissions, and essentially all of the nitrogen and phosphorus emissions in rum production operations.
2. The second most important wastewater source on a flow volume basis is the boiler and cooling water blowdown/fermenter washdown stream, the flow component of which is derived primarily from the blowdown streams, and the organic component of which is derived from the fermenter washdowns.

It is apparent from the preceding that the slops stream constitutes the major wastewater management problem in this industry.

TABLE 1. WASTEWATER GENERATION IN RUM PRODUCTION

Waste Parameter or Constituent	Total Facility Waste Generation per Proof Gallon	% Contribution by Type of Waste Stream Slops Stream	Barrel Washings	Boiler/Cooling Water & Fermenter Washdown	Water Treatment & Analytical Lab. Wastewaters
Volume	55.6 l (14.7 gal)	66%	5%	26%	3%
COD	3.0 kg (6.6 lb)	98%	1%	1%	--
BOD	1.0 kg (2.3 lb)	99%	--	1%	--
Total Solids	4.2 kg (9.2 lb)	91%	--	9%	--
Total Dissolved Solids	3.9 kg (8.6 lb)	91%	--	9%	--
Total Suspended Solids	0.25 kg (0.56 lb)	97%	--	3%	--
Total Kjehldahl Nitrogen	0.06 kg (0.14 lb)	100%	--	--	--
Total Phosphate	0.003 kg (0.007 lb)	100%	--	--	--

Source: Bacardi Corporation

The slops stream typically has the following characteristics (a detailed characterization is presented in Section VI):

1. COD - 70 to 100 gm/l
2. BOD - 20 to 60 gm/l
3. Total suspended solids - 3 to 10 gm/l
4. Total dissolved solids - 75 to 85 gm/l
5. Total nitrogen - 0.8 to 1.5 gm/l
6. Total phosphorus - 60 to 100 mg/l
7. Sulfate - 3 to 5 gm/l
8. pH - 4.0 to 4.7
9. Color - 100,000 units
10. Temperature - 80 to 90°C

The principal factors associated with the magnitude and variation of these characteristics are: the variable sugar and ash contents of the molasses, which itself is a byproduct of sugar production; and the amount of acidification (H_2SO_4) of the molasses-water mixture to obtain an optimal pH level for the fermentation. The slop is rich in nutrient materials having value as cattle feed extenders and soil supplements. From a biological treatment perspective, the slops stream typically contains a deficiency of nitrogen and phosphorus, and most of the volatile suspended solids component of the slops stream is derived from the yeast crop produced in the fermentation.

In recognition of the significance of the slops stream as a wastewater management problem for the rum industry, the present study was initiated to examine the efficacy of an anaerobic biological treatment system as the initial processing step in the treatment of slops. There are several basic reasons for selecting anaerobic biological treatment for examination, the foremost of which are:

1. The organic content of the slops (70 to 100 gm/l COD) is well in excess of the point where the oxygen demand in an aerobic system exceeds economically attainable rates of oxygen transfer.
2. The anaerobic fermentation process is characterized by low kinetic reaction constants and biomass yields, the intramolecular breakdown

of complex organic compounds, and the production of methane gas as a byproduct.

3. Sludges produced in a properly functioning anaerobic fermenter are generally more easily dewatered than are aerobic sludges, and a lesser mass of waste solids is produced per unit volume of wastewater treated in the anaerobic process than in the aerobic process.

The anaerobic biological treatment of rum distillery slops has been examined by a number of researchers prior to the present investigation. The studies conducted thus far have been done on a case-by-case basis, and while having demonstrated the amenability of slops to anaerobic fermentation, prior effort hasn't led to the establishment of a generally applicable kinetic model for predicting the performance of the anaerobic fermentation process as a function of the design variables. Additionally, the applicability of the anaerobic contact process (incorporating the recycle of biomass from a solids separator back into the anaerobic fermentation as a means of maintaining greater biomass concentrations and greater sludge ages at reduced fermenter hydraulic residence times) has not yet been examined for treatment of rum distillery slops, although its applicability for treatment of other concentrated organic streams is well documented.

STUDY ORGANIZATION

The study was initiated in May 1972 with the development of a laboratory facility and pilot plant site at the Palo Seco, Puerto Rico distillery of Bacardi Corporation. During the latter half of 1972:

1. Two bench-scale anaerobic contact process units were designed and made operational
2. Pilot plant facilities and instrumentation were leased from the University of Puerto Rico, and refurbished to incorporate a separate clarifier/thickener for use (in an anaerobic contact process configuration) in the present study
3. Work plans and preliminary reports for the investigations were developed
4. Seed culturing techniques were established and seed cultures were developed for use in the startup of each of the three units

Biological treatability studies were initiated with the startup of the two bench-scale systems in February 1973, and the pilot plant in April 1973. The bench-scale units were operated over a nine-month period under a wide range of hydraulic and organic loading rates, to develop the information necessary to establish the process kinetics (Monod kinetics) describing the anaerobic biological treatment of slops in the anaerobic contact process. The pilot plant operations were designed to develop the information necessary to verify the process kinetic parameters at the pilot-scale, and to investigate the settling characteristics of mixed liquor solids suspensions.

During the last month of pilot plant operations (December 1973), the pilot plant was modified to permit evaluation of the settling characteristics of the mixed liquor suspended solids under a wide range of surface loading rates and sludge recycle rates, both with and without polymer pre-conditioning of the mixed liquor solids suspension; the results of this terminal activity in the study were used to develop the design criteria for the gravity solids separation components of the full-scale application.

SECTION IV

DEVELOPMENT OF TECHNICAL PROGRAM

Consideration was given to several basic factors in the development of the technical program: the historical experience in the treatment of rum distillery slops by the anaerobic digestion process; prior experience in the application of the anaerobic contact process for industrial wastewater treatment; and the availability of a process kinetic model for description of process performance under steady-state conditions. The overriding interests in the process kinetic model were to: (1) have a tool to guide the experimental design and decision-making in the course of data acquisition; and (2) to provide a framework for information transfer in the design of full-scale rum distillery slops treatment systems incorporating the anaerobic contact process.

CURRENT STATUS OF TECHNOLOGY

Anaerobic Digestion of Rum Distillery Slops

Much of the experience on the treatment of spent molasses wastes by the anaerobic digestion process has been summarized in an early (1959) review by Pettet _et_ _al_(1) and in a later review by Hiatt _et_ _al_(2). Two factors reported by Pettet _et_ _al_(1) as interfering in the anaerobic digestion of rum distillery slops are the high sulfate concentrations in the slops stream, and the production of volatile acids in high concentrations during the anaerobic digestion. To circumvent the former concern (sulfide inhibition as a result of sulfate reduction to sulfide) and the general problem of high salinity concentrations in slops, much of the research to date has been conducted using diluted spent molasses (slops) waste feed streams. Stander(3) was able to obtain a 66 percent removal of organic carbon from a molasses slop diluted with an equal volume of water at a hydraulic residence time of 3.75 days, an organic loading of 8.8 kg volatile matter/day/cu m (0.55 lb volatile matter/day/cu ft), and at 33°C. In this study a repeated reinoculation technique, equivalent to a form of solids recycle, was necessary to sustain the fermentation. Hiatt _et_ _al_(2), reported that a stable fermentation was possible with a diluted slops stream (65 percent of full strength) at a hydraulic residence time of 8.4 days, an organic loading of 7.7 kg total (unfiltered sample) COD/day/cu m (0.48 lb total COD/day/cu ft), and at a temperature of 35°C. A 71-percent removal of total COD was obtained. These researchers also reported that a sustained fermentation of full strength slops was obtained at a hydraulic residence time of 16 days, and at an organic loading rate of 5.9 kg total COD/day/cu m (0.37 lb total COD/day/cu ft), with a resultant COD removal of 70 percent. After initial stabilization of the digesters, the gas produced contained about 65 percent methane, and the methane in the gas production represented about 69 percent of the COD removal. These researchers suggested that gas scrubbing to remove hydrogen sulfide and subsequent recirculation through the fermenter would be an effective control action to relieve inhibition due to sulfide ion.

The observations of Hiatt _et al_(2) that high volumetric loadings are possible if the waste is diluted confirmed earlier work on the anaerobic digestion of rum distillery slops by Radhakrishnan _et al_(4). In laboratory studies conducted at 37°C, these investigators found that the maximum possible loading with undiluted waste was 3 kg BOD/day/cu m (0.18 lb BOD/cu ft/day) at a hydraulic residence time of 11 days, whereas a maximum loading of 4.5 kg BOD/day/cu m (0.28 lb BOD/cu ft/day) was attained at a five-day hydraulic residence time with a diluted waste (60 percent of full strength). BOD removals in these studies averaged from 80 to 85 percent. The gas contained from 54 to 60 percent methane, and more than one percent hydrogen sulfide. In evaluation of their results, the authors reported that volatile acids levels in excess of 2,000 mg/l (as acetic acid) did not impair the fermentation, providing that a sufficiency of alkalinity was present to maintain the pH of the system in excess of 7.0. Additionally, sulfide concentrations as high as 240 mg/l were not found to be inhibitory to the digestion of undiluted wastes.

The preceding examples of prior experience with the anaerobic digestion of spent molasses wastes are representative of the state-of-the-art to date in that:

1. The efficacy of anaerobic digestion of this type of waste has been established;

2. Two types of toxic/inhibitory factors are commonly identified in the applications: (a) high feed stream sulfate concentrations; and (b) high volatile acid concentrations;
3. The high sulfate concentrations have been dealt with in several investigations by reduction of the strength of the waste stream to levels where its effect was deemed marginal
4. The high volatile acids concentrations anticipated in the digestion process have been dealt with either by dilution of the raw waste or by control of the rate of loading of undiluted wastes such that the rate of alkalinity production was sufficient to balance the rate of volatile acid generation.

Unfortunately, due to the inconsistent frameworks used in reporting information derived in prior investigations, it is not possible to characterize the anaerobic digestions achieved in prior studies and to evaluate the effectiveness the operational controls used on a rational kinetics basis, or even on a solids balance basis. Without such information, it is impossible to interpret the significance of empirical measures such as volumetric organic loading rate or hydraulic residence time in terms of the viable biomass inventory associated with the observed performance; nor is it possible, in the absence of information on such parameters as mixed liquor solids concentrations to ascertain, or even imply, the nature of the solids handling problems to be dealt with in the design and operation of a full-scale application.

The absence of a rational basis to characterize the prior work raises further questions with respect to evaluation of the operational strategies used to control toxic substances in the fermentation of spent molasses wastes. That is, there are two generic classes of toxic substances, those which can serve as substrates after acclimation and those which cannot serve as substrates. To compensate for toxicants in the former class it is necessary to decrease the mass loading rate of biodegradable matter per unit mass of viable organisms by one or more of the following strategies (5):

1. Reduction of the mass loading rate of biodegradable matter, holding the viable biomass constant;

2. Increasing the viable biomass while holding the mass loading rate of biodegradable matter constant

In the latter category of toxicants, and depending on the type of inhibition (competitive or noncompetitive), the alternative strategies are:

1. Reduce the concentration of the toxicant to a level below which its effect is marginal or neutralized by other chemical/biological interactions
2. Neutralize the effect of the toxicant by the addition of chemical agents
3. Reduce/eliminate the concentration of the toxicant in the waste stream prior to anaerobic treatment (i.e., source control)

The volatile acids generated in the anaerobic fermentation are indicative of a toxicant in the former category in that volatile acids can serve as a substrate, but judicious operation is required. Unfortunately one cannot ascertain from the literature the ranges of rational loading rates (e.g. lb volatile matter per lb solids per day) in which the volatile acids concentration can be stabilized in the application.

Several authors (2,6,7) discuss the ionic content of spent molasses wastes, the origin of which can be related to the "salt pickup" in the production of molasses, the chemical conditioning of the molasses mixture prior to fermentation, the "pickup" of metal components from equipment in the distillation of the ferment, etc. While sulfides are classifiable as a toxicant which cannot serve as a substrate in anaerobic fermentation, there has been little examination in prior investigations of the chemical interactions between the sulfide ion generated in the reduction of sulfate in the anaerobic fermentation of spent molasses wastes, and the heavy metal ion inventory found in these wastes.

Given the absence of a transferable information base for the design and application of a full-scale anaerobic treatment system for rum distillery slops, the overriding priority of the present study was deemed to be the development of a rational kinetic basis for unifying the understanding of the performance characteristics of anaerobic fermentation systems in the application. Given the importance of solids recycle in biological treatment systems, attention was directed to the anaerobic contact process.

Anaerobic Contact Process

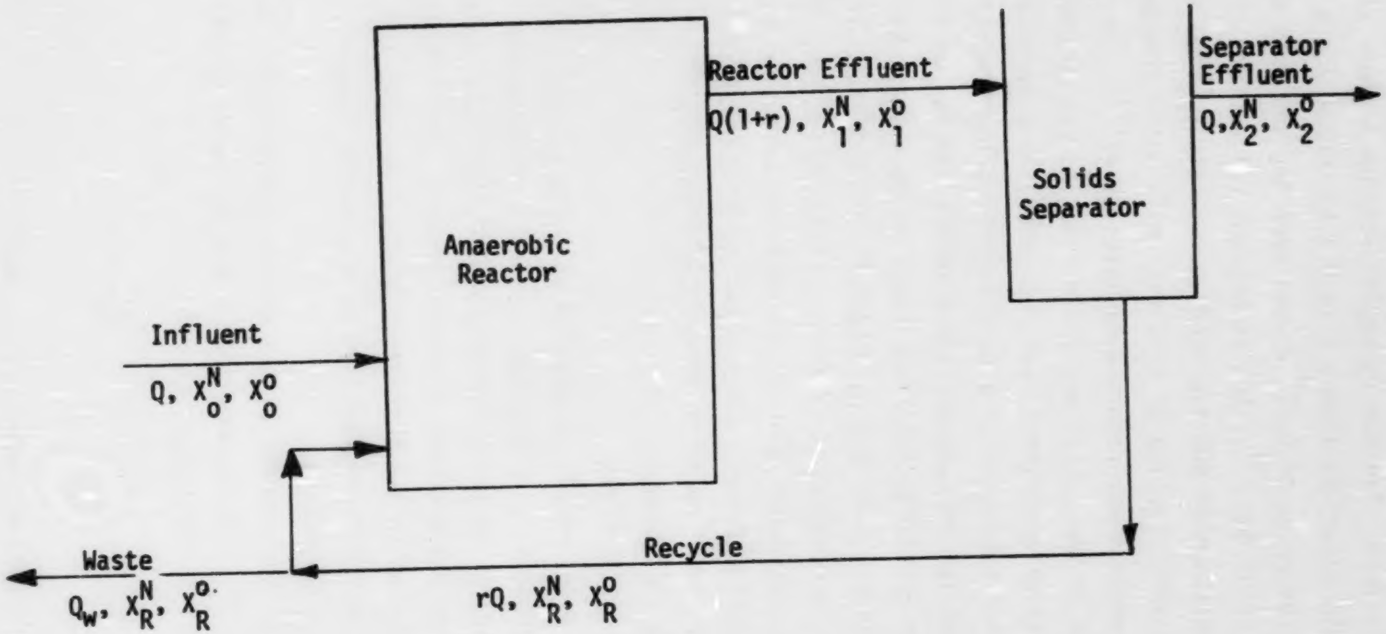
The anaerobic contact process (Figure 1) basically consists of two unit processes, an anaerobic fermenter and a solids separator (either gravity or mechanical), with provisions for recycle of concentrated biomass from the separator to the fermenter and complete mixing of the fermenter. Heating of the fermenter contents is provided in those cases where higher throughput rates (resulting in hydraulic residence time of hours rather than days) do not preclude its economic feasibility, or when sufficient methane is produced in the fermentation to satisfy heating requirements.

Fundamentally, the anaerobic contact process as originally described and documented (8,9) differs from anaerobic sludge digestion in that biomass recycle and relatively low substrate concentrations are used. The literature contains reports on the use of the anaerobic contact process for the treatment of cannery, packinghouse, brewery, distillery, fatty-acid, wood fiber, and synthetic-milk wastewaters.

In evaluation of the literature, it was found that an anaerobic process was originally classified (on the basis of applications experience) to be of the contact type if it met the following criteria(9): (1) an influent concentration of 4,000 mg/l BOD or less; (2) use of sludge recycle; and (3) a liquid detention time of four days or less based on influent flow. Thus, the application of the anaerobic contact process for the treatment of rum distillery slops, in the present study, has represented an initial effort toward establishing the feasibility of the anaerobic contact process for treatment of high-strength wastes containing BOD concentrations in excess of 4,000 mg/l.

The slaughterhouse application represents the most successful and detailed study of the anaerobic contact process to date. The initial stages of this study were reported by Fuller(10). Subsequent work was presented by Schroepfer *et al*(11), and Schroepfer and Ziemke(12, 13). The laboratory and pilot-plant work conducted by these investigators resulted in the construction and operation of a full-scale facility(14). During the series of investigations it was determined that vacuum degasification preceding gravity sedimentation was the most suitable method for obtaining sufficient solids concentration to permit continuous

FIGURE 1. SCHEMATIC OF GENERAL TREATMENT SYSTEM


Anaerobic Reactor
Reactor Effluent
$Q(1+r), X_1^N, X_1^o$
Solids Separator
Separator Effluent
Q, X_2^N, X_2^o
Influent
Q, X_o^N, X_o^o
Recycle
rQ, X_R^N, X_R^o
Waste
Q_W, X_R^N, X_R^o

solids recycle. Reactor hydraulic residence times as low as 2.3 hours and loadings varying from 0.6 to 5.3 kg BOD/day/cu m were used successfully. The range of θ_T/C^o values reported for this application was from 12 to 45 days(9). BOD reductions ranged from 70 to 97 percent, and raw wastewater BOD averaged approximately 1,400 mg/l.

The investigation of synthetic-milk waste treatment by the anaerobic contact process was reported by Gates et al(9). The investigation was conducted at the laboratory-scale using an upflow anaerobic contact unit (settler positioned over fermenter) in a vertical configuration. The settler consisted of a baffle-type upflow separator which showed good potential for accomplishing solids separation in the process. Fermenter hydraulic residence times as low as six hours and loadings varying from 0.8 to 5.4 kg BOD/day/cu m (0.05 to 0.34 lb/day/cu ft) were used successfully. BOD reductions varied from 30 percent at a hydraulic residence time of six hours to levels in excess of 75 percent at hydraulic residence times equal to or greater than 12 hours. θ_T/C^o values ranging from 10 to 28 days were reported(9). The influent BOD concentration averaged 1,200 mg/l.

PROCESS KINETIC MODEL

The Monod model, as adapted by Gates et al(9) to the anaerobic contact process, was selected for use in the present study because of its ability to describe the performance (effluent substrate concentration) produced in the anaerobic contact process as a function of the hydraulic residence time in the fermenter, biomass recycle, and kinetic constants specific to the biodegradation of a given wastewater.

Basic Relationships

In the development of the growth kinetic model based on the work of Monod(15) it is assumed that gross bacterial growth is always exponential in character, that is:

$$\frac{dX^o}{d\theta} = k \qquad (1)$$

where:

X^o = organism concentration, mg/l

θ = time, and

k = specific growth rate, time^{-1}

The expression for net growth then becomes:

$$\frac{dX^o}{d\theta} = (k - K^D)X^o \tag{2}$$

where K^D is specific autodestruction rate.

The relationship determined by Monod(15) for the growth-rate constant is:

$$k = \frac{k^m X^N}{K+X^N} \tag{3}$$

where:

k^m = maximum specific growth rate, time^{-1}
X^N = concentration of controlling substrate, mg/l, and
K = substrate concentration when $k = k_m/2$, mg/l

Combining Equations 2 and 3 gives the overall expression for the change in organism concentration with time; that is:

$$\frac{dX^o}{d\theta} = \left[\left(\frac{k^m X^N}{K+X^N}\right) - K^D\right] X^o \tag{4}$$

For wastewater treatment processes, however, the major interest is in the change of substrate concentration with time rather than a terminal interest in the growth characteristics and capabilities of the bacteria. The relationship between organism concentration change with time and substrate concentration change with time can be expressed as follows:

$$- \frac{dX^o}{d\theta} = Y^o \frac{dX^N}{d\theta} \tag{5}$$

where Y^o is mass of organisms produced per mass of substrate descrease (mg/mg).

Combining Equations 1, 3 and 5 yields: -

$$\frac{dX^N}{d\theta} = -\left(\frac{1}{Y^o}\right)\left(\frac{k^m X^N}{K+X^N}\right) X^o \tag{6}$$

which is the working relationship describing the rate of substrate consumption when the Equation of Monod(15) is used.

When Equations 1, 2, 3, and 6 are applied to a continuous flow, completely-mixed reactor at steady rate with solids recycle (as exemplified in Figure 1), the equations required to describe the characteristics of such a reactor can be derived from substrate and solids balances around the anaerobic contact system. The resultant equations are presented below, and the physical significance of the symbology used is illustrated in the process flow sheet of Figure 1:

$$X_1^o = \frac{Y^o(X_o^N - X_1^N)}{K^D\theta_T + C^o} \qquad (7)$$

$$X_1^N = \frac{K\ (K^D\theta_T + C^o)}{k^m\theta_T - (K^D\theta_T + C^o)} \qquad (8)$$

$$k = K^D + C^o/\theta_T \qquad (9)$$

$$C^o = X_w^o/X_1^o \qquad (10)$$

where:

X_1^N = effluent organism concentration, mg/l

X_o^N = influent substrate concentration, mg/l

X_1 = effluent substrate concentration, mg/l

K^D = specific autodestruction rate, time^{-1}

θ_T = theoretical hydraulic detention time,

C^o = recycle factor (unitless), and

X_w^o = mass of organisms leaving system per unit time/total flow leaving system per unit time, mg/l.

In general, the value of X_w^o is the weighted average of the concentration of organisms leaving the solids separator in the effluent and in any sludge that would be wasted. The value of C^o is zero when all the organisms are recycled and one when no recycle is used.

The above equations assume that the controlling substrate is non-settleable, thus, the substrate level in the system's effluent is the same as that in the reactor. The equations can be modified to consider a settleable substrate, as was observed to be the case in both the bench and pilot systems used in the present investigation. The modified equations as used herein were developed to account for the loss of substrate and biomass from the anaerobic contact process in both sampling and sludge wastage streams. The modified forms of Equations 7 and 10 are as follows (no modifications were required to Equations 8 and 9):

$$X_1^o = \frac{Y^o \left[X_o^N - X_1^N \left[1 - r\,(C'' - 1) \right] \right]}{K^D \theta_T + C^o} \tag{11}$$

and:

$$C^o = \frac{Q_w X_R^o + (Q - Q_s - Q_w)\, X_2^o}{Q X_1^o} + \frac{Q_s}{Q} \tag{12}$$

where:

X_R^N = recycle stream substrate concentration, mg/l

X_R^o = recycle stream (and sludge wastage stream) organism concentration, mg/l

Q_s = sample flow rate (average), l/time (not shown in Figure 1),

Q_w = sludge wastage flow rate, l/time

C'' = degree of concentration of substrate in the solids separator ($= X_R^N / X_1^N$),

r = fraction of influent recycled.

Determination of Kinetic Constants

For purposes of determining the kinetic constants, Y^o, K^D, k^m, and K of the model, engineering-type parameters such as soluble (filterable)

COD have been used to represent the rate-controlling (biodegradable) substrate concentration (X_1^N) and volatile suspended solids (VSS) as a parameter representing the viable biomass concentration (X_1^o) in the above equations. The determination of the process kinetic constants requires process evaluation (flow, solids and substrate balances) over a range of variation of θ_T/C^o values (biomass retention times) of one-half to one order of magnitude.

The kinetic constants are determined by using linear forms of Equations 11 and 8 (Figure 2) to analyze the experimental points obtained in the balances. From Equation 11:

$$\frac{X_o^N - X_1^N \left[1-r\,(C''-1)\right]}{C^o X_1^o} = \frac{1}{Y^o} + \frac{K^D}{Y^o}\left(\frac{\theta_T}{C^o}\right) \qquad (13)$$

From Equation 8:

$$\frac{\theta_T}{K^D\theta_T + C^o} = \frac{1}{k^m} + \frac{K}{k^m}\left(\frac{1}{X_1^N}\right) \qquad (14)$$

A plot of Equation 13 is presented in Figure 2a, in which the intercept is the inverse of the value of Y^o, and the slope represents the ratio K^D/Y^o, for the best-fit line of the set of experimental data used. Presented in Figure 2b is a plot of Equation 14; the intercept of the best-fit line in this figure is the inverse of the value of k^m and the slope is equal to K/k^m. In consideration of the best-fit lines used in determining the kinetic constants, it is apparent that from five to ten well-spaced experimental points are sufficient to define the process with the Monod model in any application.

Operational Relationships

Equation 8 can be rewritten to formulate an operational relationship between the effluent substrate concentration, X_1^N, of the anaerobic

FIGURE 2. DETERMINATION OF KINETIC CONSTANTS

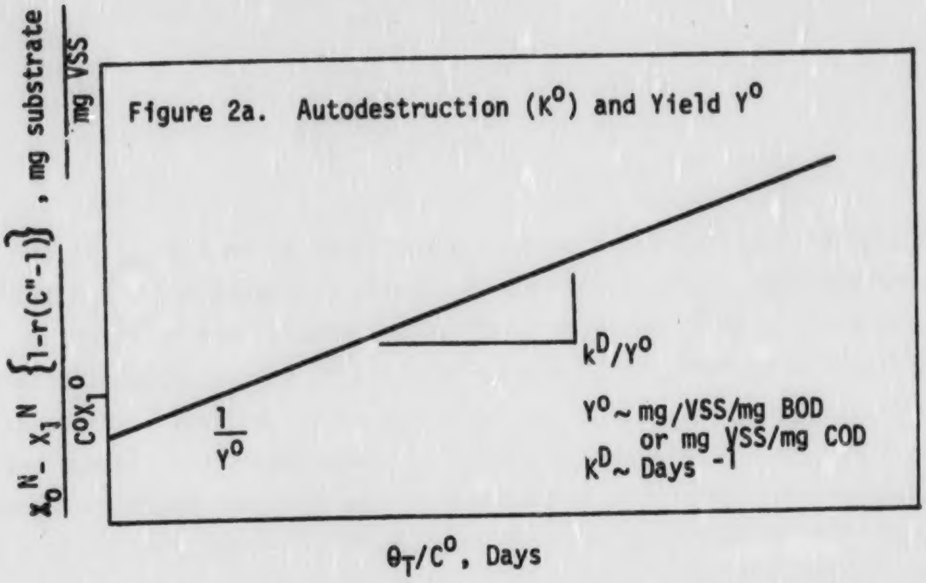


Figure 2a. Autodestruction (K^o) and Yield Y^o

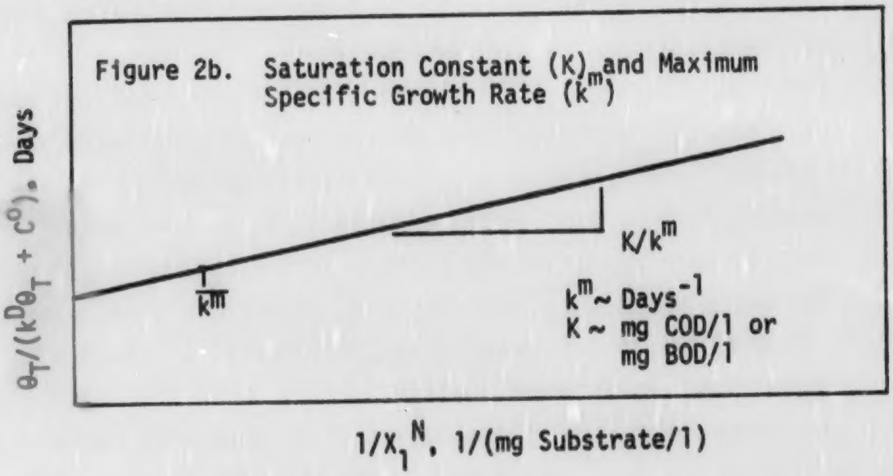


Figure 2b. Saturation Constant (K) and Maximum Specific Growth Rate (k^m)

contact process, and the basic design parameters, θ_T (hydraulic residence time) and C^o (recycle factor), as follows:

$$\frac{\theta_T}{C^o} = \frac{K + X_1^N}{(k^m - K^D)X_1^N - KK^D} \qquad (15)$$

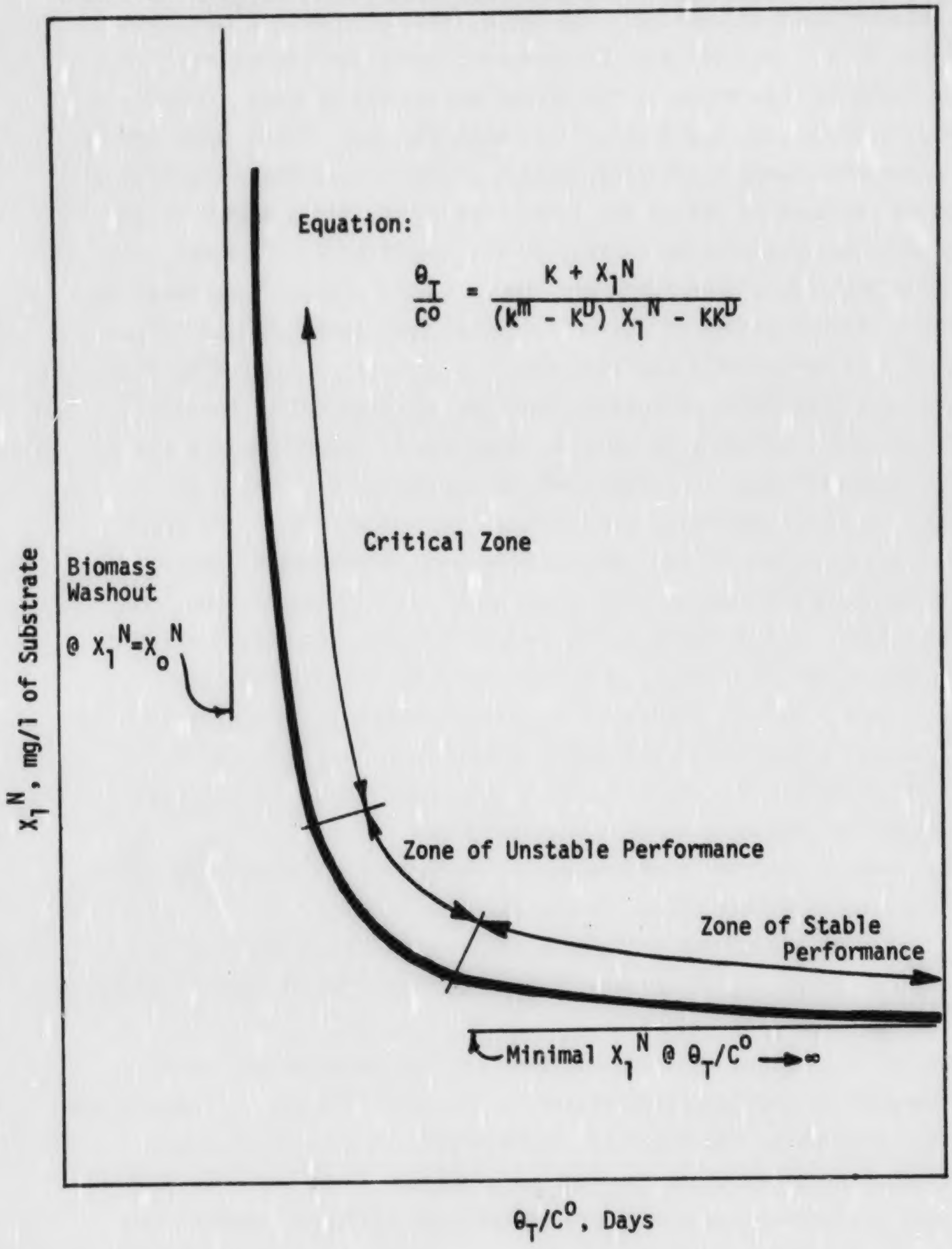
A plot of Equation 15 is presented in Figure 3, from which it is apparent that, for a given set of kinetic constants and a specified effluent quality, the ratio θ_T/C^o is specified. The relationship defined by Equation 15 is a hyperbola which becomes asymptotic to an ordinate value equal to $K^D K/(k^m - K^D)$ and to an abscissa value equal to $1/(k^m - K^D)$. The value of $K^D K/(k^m - K^D)$ represents the minimum substrate concentration which can be realized in the effluent and is obtained when C^o is zero (i.e., 100 percent recycle of biomass) or θ_T is infinity. The value of $1/(k^m - K^D)$ is approached as the effluent substrate concentration (X_1^N) approaches the influent concentration (X_o^N), i.e., when biomass washout occurs.

Several basic premises of the kinetic formulation for the anaerobic contact process are embodied in this expression:

1. Effluent quality as predicted by Equation 15 is directly a function of the viable biomass in the fermenter.
2. Effluent quality is not a function of influent quality; the only way that the biological process can respond to changes in the influent quality (or more appropriately, to changes in the mass loading rate of biodegradable matter), is by an increase or decrease in the inventory of viable biomass in the system.
3. For any given θ_T, an increase in C^o results in a decrease in effluent quality; a decrease in the value of C^o results in an improvement in effluent quality.

The basic objective of the design of an anaerobic contact system is to achieve the θ_T/C^o ratio as specified by the desired effluent quality and the determined kinetic constants for the least cost. The

FIGURE 3. UNIFIED OPERATIONAL RELATIONSHIP FOR ANAEROBIC CONTACT PROCESS



costs associated with θ_T are the tankage cost, the cost of mixing, and the optional cost of heating. The costs associated with C^o are those for the solids separator and the necessary pumps and piping, and the power costs for operation of the separator and for pumping. Thus, a consideration of the costs associated with the size of the fermenter and separator, pumps and piping, mixing and power and other operational costs can be used to define the least cost region for a design of an anaerobic constant process to realize the specified θ_T/C^o value.

The problems of operation are somewhat more arduous than those of design. The net result of design is almost inevitably the specification of a fermenter of fixed size (which provides the desired θ_T based on a design flow rate), a solids separator of fixed solids handling capacity, and a capacity to vary the quantity of underflow from the solids separator that is pumped back to the fermenter. Thus, in effect, the only controllable operational variable is C^o. The major factors which affect C^o are the effectiveness of the solids separator in controlling the biomass concentration in the liquid effluent from the separator, and the capacity to recycle biomass from the clarifier. After the design of a solids separator has been established, its effectiveness can be controlled only by the recycle rate and/or the addition of primary coagulants or coagulant aids (i.e. polymers).

Three regions of operation can be defined with the operational curve for the anaerobic contact process (Figure 3):

1. A zone of "stable" performance, in which a unit change in the θ_T/C^o results in little change in the effluent quality.
2. A zone of transition or "unstable" performance, in which a unit change in the θ_T/C^o value will generate significant change in the effluent quality.
3. A critical zone, in which effluent quality deteriorates rapidly.

Because as previously discussed, C^o is typically the only operational variable available, the anaerobic contact process has its greatest operational sensitivity in the "unstable" zone. Thus, when the desired effluent quality is based on a pretreatment or effluent standard (the most likely case), the objective of the operation becomes primarily that of ensuring that C^o is equal or less than the design value at all times. However, if the actual θ_T is variable, that is, if the actual flow rate deviates widely from the design flow value, then the objective of operation is to vary C^o such that at all times, the ratio θ_T/C^o is equal to or greater than that necessary to provide the desired effluent quality.

RATIONALE AND EXPERIMENTAL STRATEGY

The experimental program was organized to deal with the following types of information needs:

1. Delineation of criteria for startup of a methanogenic fermentation.
2. Delineation of the process performance characteristics of the anaerobic contact process in this application within the context of Monod kinetics.
3. Documentation of the performance characteristics of alternative solids separation and handling unit processes.
4. Development, from the above, of design criteria for a full-scale application.
5. Economic evaluation of the full-scale application on a cost-performance basis.

In consideration of these needs, the following sequence of steps were taken in the conduct of the experimental program:

1. A process kinetic model of the anaerobic contact model (described herein) was selected as the basis for:
    a. providing a unifying framework for integrating the use of biomass recycle to overcome the low growth rates of the methanogenic bacteria, such that the resultant process descriptions are transferable from one case to the next.
    b. experimental design, i.e., for the design of experimental facilities, selection of operating conditions, and for the continuous updating of experimental conditions being examined as process response information was developed.

2. Bench-scale anaerobic contact systems (incorporating 20-liter anaerobic fermenters) were developed and operated over a range of hydraulic loading rates sufficient to permit definition of the kinetic constants of the Monod Model.
3. The kinetic constants were used to develop an operational relationship for predicting pilot plant performance as a function of θ_T (hydraulic residence time) and C^o (recycle factor), and the relationship was used to select pilot plant operating conditions to be examined.
4. The pilot-scale anaerobic contact system (incorporating a 1,890 liter anaerobic fermenter) was operated at the selected operating conditions to provide a data base for confirming the kinetic relationships.
5. Upon completion of the process kinetic characterization, the pilot-plant system was modified to permit operation of the clarifier over a wide range of solids loading rates, using the mixed liquor suspended solids in the pilot plant as a recyclable feed stream,
6. Laboratory studies were conducted prior to the pilot-scale settleability testing to guide the selection of a coagulant aid and to develop dosing criteria with the selected coagulant aid.

SECTION V

EXPERIMENTAL APPARATUS AND PROCEDURES

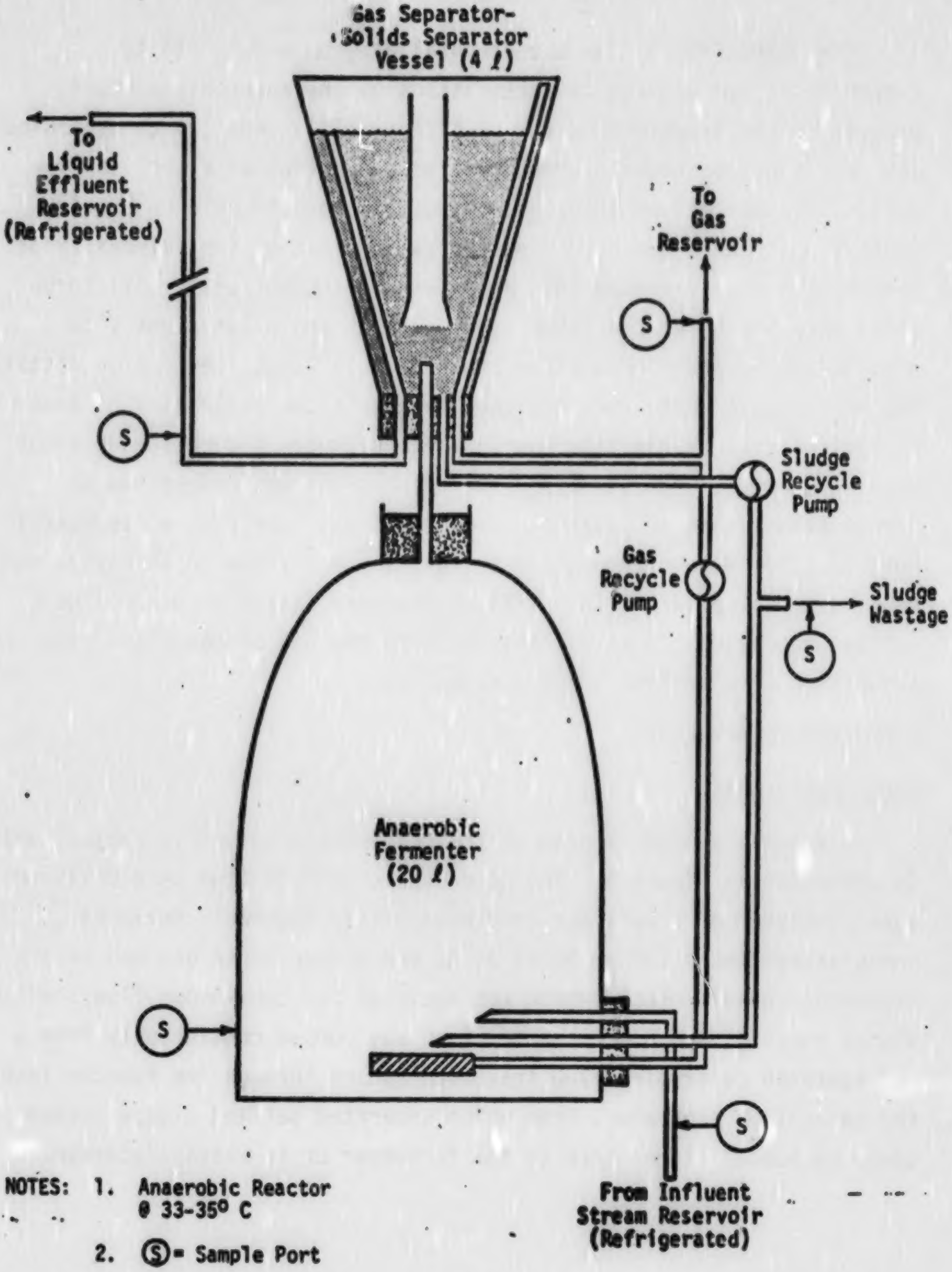
The objectives of the experimental program were: (1) to document the performance characteristics of the anaerobic contact process in the treatment of rum distillery slops, and (2) to determine the clarification capacity required for separation of mixed liquor solids. To accomplish these objectives, two bench-scale anaerobic contact units were designed, and the pilot plant of the University of Puerto Rico was converted into an anaerobic contact unit. All three units were set up at the laboratory facility and pilot plant site provided by Bacardi Corporation at their Palo Seco, Puerto Rico distillery. The bench-scale units were designed to permit the evaluation of anaerobic biological process kinetics over a range of hydraulic residence times varying from five to 200 days. The pilot plant was redesigned to permit examination of anaerobic biological response over a necessarily more limited range of hydraulic residence times (five to 50 days), and to permit evaluation of the settling characteristics of mixed liquor solids suspensions, both without and with the use of coagulant aids, as a function of clarifier loading rate.

EXPERIMENTAL APPARATUS

Bench-Scale Units

A schematic flow diagram of the bench-scale anaerobic contact unit is presented in Figure 4. The bench-scale systems (two were utilized) each consisted of a 20-liter completely-mixed anaerobic fermenter, temperature-controlled at 33 to 36°C, and a four-liter gas and solids separator vessel. Each system was equipped for continuous flow feeding, sludge recycle, and gas recycle. Feed was pumped continuously from a refrigerated reservoir; flow traversed upward through the reactor into the gas/solids separator, from which separated gas and sludge streams could be pumped either back to the fermenter or to wastage/storage.

FIGURE 4. BENCH-SCALE ANAEROBIC CONTACT UNIT


Gas Separator-
Solids Separator
Vessel (4 ℓ)
To
Liquid
Effluent
Reservoir
(Refrigerated)
To
Gas
Reservoir
S
S
Sludge
Recycle
Pump
Gas
Recycle
Pump
Sludge
Wastage
S
Anaerobic
Fermenter
(20 ℓ)
S
S
From Influent
Stream Reservoir
(Refrigerated)

NOTES: 1. Anaerobic Reactor @ 33-35° C

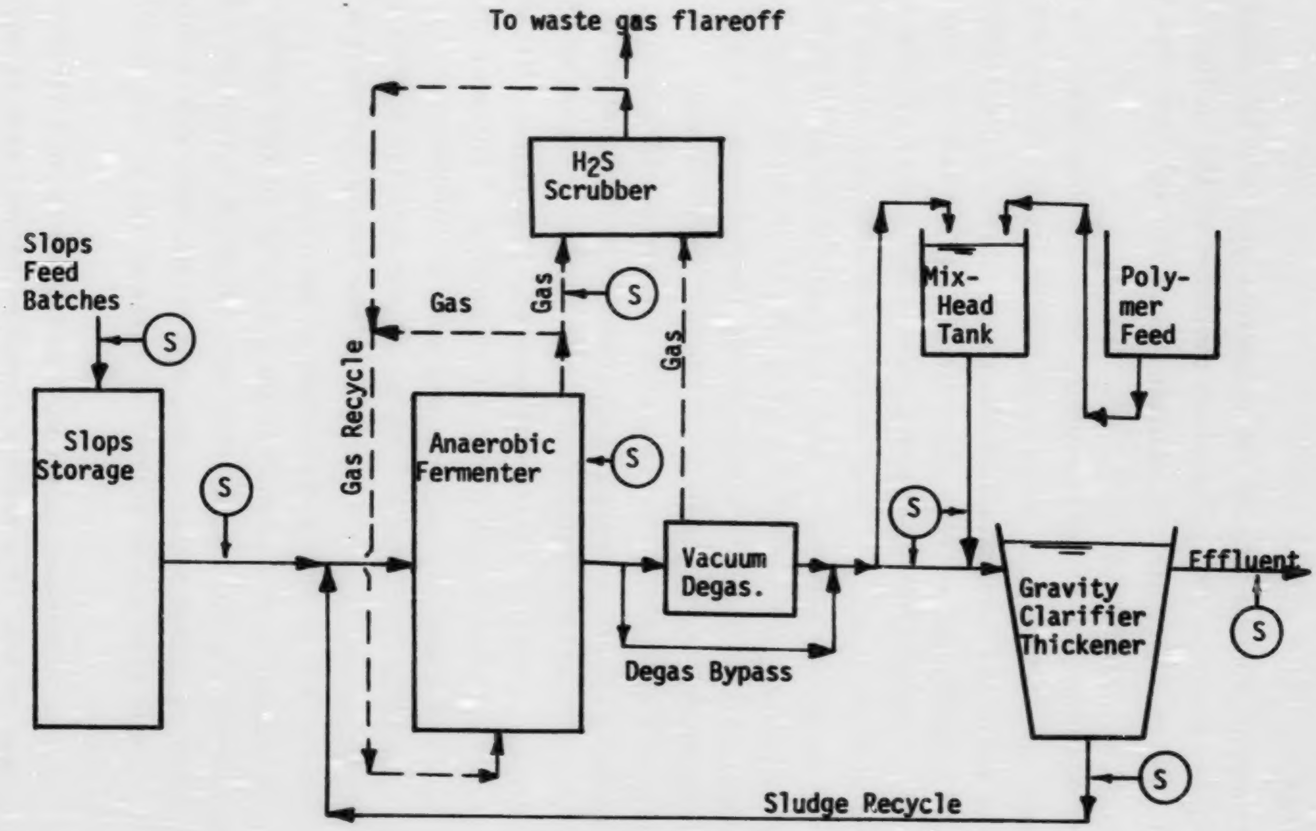
2. Ⓢ = Sample Port

Pilot Plant

A schematic diagram showing the components of the pilot plant is presented in Figure 5. The pilot plant consisted of the following:

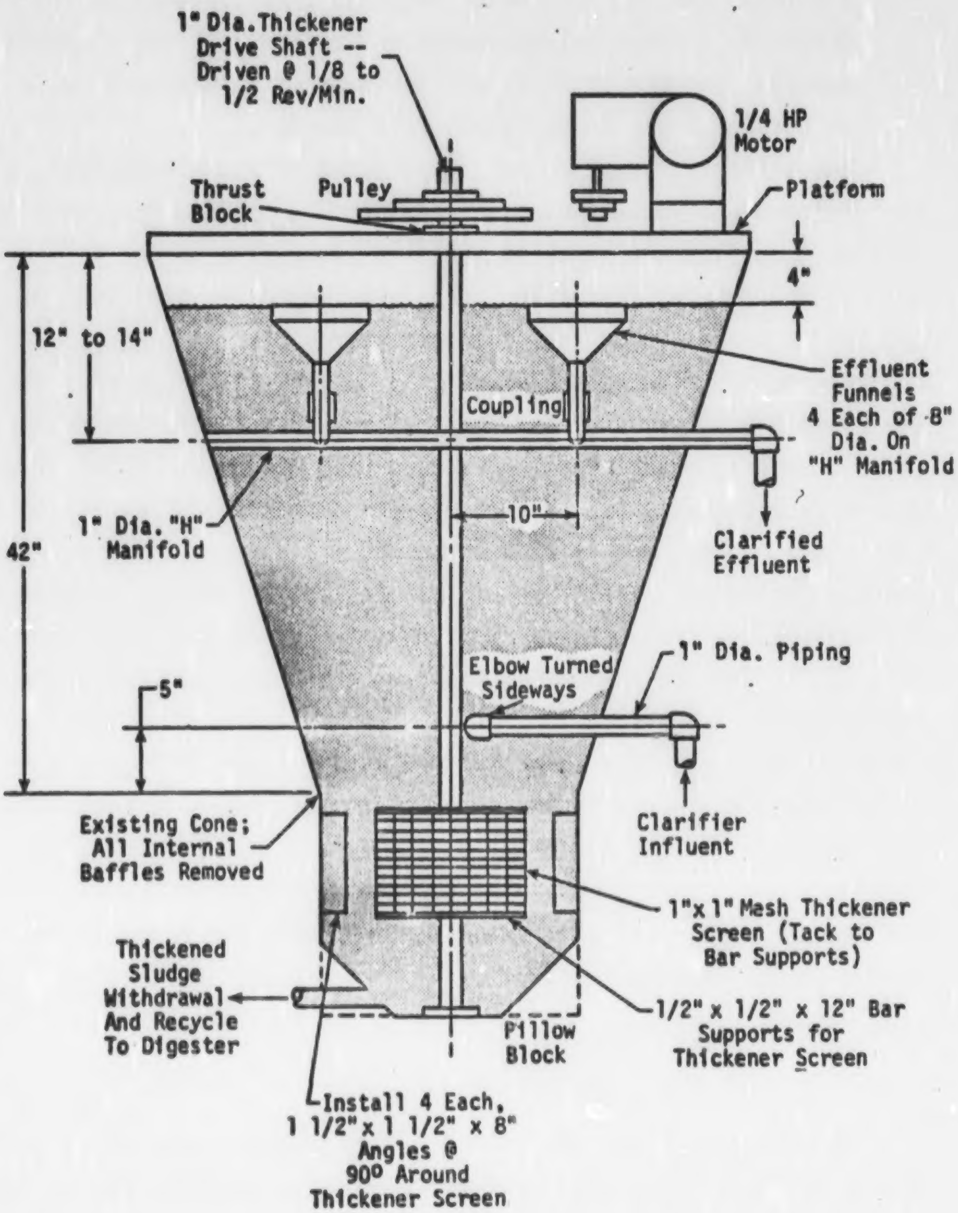
1. A 1,130-liter (300-gallon) insulated slops storage tank used to hold feed slops batches.
2. A 1,890-liter (500-gallon) insulated anaerobic fermenter, equipped with: an electric heat tape system, thermostatically controlled to maintain the temperature of the mixed liquor in the fermenter at 33 to 36°C; gas recycle at a rate of 0.15 standard cu m/min/cu m of mixed liquor (20 SCFM/1,000 gallons); and liquid recycle at a rate of 0.06 cu m/min/cu m of mixed liquor (60 gpm/1,000 gallons)
3. Hydrogen sulfide scrubbers (two), each consisting of cylindrical tanks, 0.1 m diameter by 0.6 m in height (4-in diameter by 2-ft height), filled with steel wool and designed for removal of approximately 1/2 kg elemental sulfur per kg steel wool
4. A vacuum degasifier, consisting of a cylindrical tank 0.15 m diameter by 0.6 m in height (6-in diameter by 2-ft height), for degasification of the mixed liquor flow at a vacuum of 15 mm Hg.
5. A gravity clarifier/thickener (Figure 6), constructed from the internal conical element of the pilot plant as furnished; the key features of the clarifier/thickener were as follows:
    a. the conical (clarifier) sector of the unit has a cross-sectional area of one sq m (10.7 sq ft) at the liquid surface and 0.21 sq m (2.2 sq ft) at the bottom of the cone.
    b. effluent is withdrawn from the clarifier in four effluent funnels, presenting a total weir length of 2.57 m (8.4 ft).
    c. the thickener sector of the unit is equipped with a thickener screen that can be rotated at rates of 1/8 to 1/2 revolution/minute.
    d. sludge withdrawal is accomplished by means of a variable speed pump capable of a maximum sludge withdrawal rate of 985 liters/day (260 gallons/day).

FIGURE 5. PILOT PLANT PROCESS SCHEMATIC


To waste gas flareoff
H2S Scrubber
Slops Feed Batches
S
Gas
Gas
Gas
Gas Recycle
Slops Storage
Anaerobic Fermenter
Vacuum Degas.
Degas Bypass
Mix-Head Tank
Poly-mer Feed
Gravity Clarifier Thickener
Effluent
Sludge Recycle

Note: Ⓢ = Sampling Point

FIGURE 6. PILOT PLANT CLARIFIER/THICKENER


1" Dia. Thickener Drive Shaft -- Driven @ 1/8 to 1/2 Rev/Min.
1/4 HP Motor
Thrust Block
Pulley
Platform
4"
12" to 14"
Effluent Funnels 4 Each of 8" Dia. On "H" Manifold
Coupling
1" Dia. "H" Manifold
10"
42"
Clarified Effluent
Elbow Turned Sideways
1" Dia. Piping
5"
Clarifier Influent
Existing Cone; All Internal Baffles Removed
1"x 1" Mesh Thickener Screen (Tack to Bar Supports)
Thickened Sludge Withdrawal And Recycle To Digester
Pillow Block
1/2" x 1/2" x 12" Bar Supports for Thickener Screen
Install 4 Each, 1 1/2" x 1 1/2" x 8" Angles @ 90° Around Thickener Screen

6. A 190-liter (50 gallon) mix-head tank and a stock polymer solution feed tank; the mix-head tank was equipped with a constant-head overflow (the overflow being recycled directly back to the anaerobic fermenter); stock polymer solution was fed into the mix-head tank and the contents of this tank could be mixed with a recycle pump at a rate of one volume every five minutes.

The entire pilot plant was constructed of stainless steel, and was equipped and instrumented to provide: (a) sludge recycle from the gravity clarifier/thickener to the slops feed line to the fermenter; and (b) complete mixing of the anaerobic fermenter utilizing recycled gas and/or mixed liquor from the fermenter.

During the first phase of pilot plant operations (evaluation of anaerobic biological process kinetics), raw slops were fed from the slops storage tank at preselected continuous flow rates, and flows of mixed liquor from the anaerobic fermenter were transferred directly to the pilot plant (i.e., the flow path through the mix-head tank was not used). During the final phase of pilot plant operations (evaluation of mixed liquor settling characteristics), all mixed liquor flows were directed to the mix-head tank, and from this point by gravity into the clarifier-thickener.

Both the bench and pilot systems were designed with the sampling ports necessary to permit characterization of liquid volumes within each component of the system, liquid and material transfer rates into, within, and from the components of systems, and gas production rates. The approximate locations of the sampling ports are shown in the schematics of Figures 4 and 5.

OPERATIONAL PROCEDURES

Process Kinetic Characterization

The biological process kinetic studies were conducted in three phases of operation: seed culture development; startup and acclimation of the bench and pilot scale units; and routine operations.

Development of Seed Cultures - Seed cultures for use in the startup of the bench and pilot systems were propagated in anaerobic vessels equipped with feed and sampling ports, and with bunsen-type gas release valves. Selected as a source of methanogenic material was the bottom mud of a brackish water inlet to San Juan Bay (Puerto Rico); this inlet has received slops discharges from a small rum distillery and cooling water discharge from an electrical power generation station for a number of years. Each seed culture vessel was filled with deoxygenated tap water at 35°C; about 40 percent of which was displaced with a $CO_2:N_2$ gas mixture; bottom mud was added in an amount of about 25 percent of the volume of the vessel; and then the vessel was "topped off" with undiluted slops that had been neutralized to pH 7.0 with sodium bicarbonate.

The seed cultures were maintained by daily batch feeding of gradually increased amounts of raw slops neutralized to pH 7.0 with sodium bicarbonate. Daily observations were made on the pH of the seed culture and the gas production and composition generated in each vessel. The first trace of methane was typically observed within two to four weeks after initiation of seed culturing and a minimum time of four months was allowed before the seed cultures were transferred to the fermenters.

Startup and Acclimation - The two bench-scale and the pilot-scale anaerobic fermenters were "seeded" with the seed cultures using the same procedure as outlined above, i.e., the fermenters were filled with water, a portion of the water volume displaced with a $CO_2:N_2$ gas mixture, the seed culture added, and neutralized slops added, in the proportions outlined above. After seeding, all three units were fed neutralized (pH 7.2) undiluted slops at flow rates allowing hydraulic residence times in excess of 30 days, to permit the buildup of MLVSS (mixed liquor volatile suspended solids) concentrations to minimal levels of 300 mg/l.

Throughout the four-month acclimation period; as well as in subsequent periods of operation, each anaerobic contact unit was monitored in accordance with the schedule presented in Table 2 and described in detail below.

TABLE 2. MONITORING/ANALYTICAL SCHEDULE

Measurement or analysis	Influent stream[1]	Fermenter contents[1]	Liquid effluent[2]	Sludge wastage	Gas reservoir
Flow rates (all)	D		D	-[4]	D
COD	-[3]	TW	TW		
BOD	-[3]				
TSS/VSS	-[3]	TW	TW	-[5]	
TKN	-[3]	BM			
NH_3-N	-[3]	BM			
Total P	-[3]	BM			
Volatile Acids	-[3]	D			
pH	-[3]	D			
Alkalinity		D			
Gas Composition					D
Temperature		D			
Sulfides		W			
Sulfate	-[3]				

Notes:

[1] Continuous flow influent stream to anaerobic fermenters.
[2] Gas separator-solids separator.
[3] Done once per batch of refrigerated feed in bench system and once per batch of stored slops in pilot plant.
[4] Measure volume/record date and time for each withdrawal.
[5] For each sludge withdrawal.

D = grab sample daily, W = grab sample weekly, BM = grab sample bimonthly; TW = grab sample three times per week.

An evaluation of the operating experience acquired during the four month period of acclimation and buildup of a viable biomass inventory in each unit led to the development of the following guidelines for startup:

1. The maximal rate of slops feed should not exceed on kg soluble COD/day/kg MLVSS if the buildup of volatile acids is to be avoided
2. The pH of the mixed liquor should be maintained in the range of 7.2 to 7.3 at all times.
3. Continuous gas and/or liquid recycle mixing of the fermenter contents should be provided at all times.

Routine Operations - After completion of acclimation, it was possible to sustain a methanogenic fermentation of undiluted and unneutralized slops stream at a pH in excess of 7.0 during operations at hydraulic residence time greater than of 20 days. Neutralization was required to maintain a mixed liquor pH in excess of 7.0 at residence times below 20 days, indicative of a differential between the required and the internally-generated buffering capacity; this differential was observed to increase with decreasing hydraulic residence times at hydraulic residence times less than 20 days (Section VI). During routine operations, neutralization was accomplished, as needed, by the addition of sodium bicarbonate directly into the bench and pilot-scale fermenters.

Monitoring and Analytical Program - A summary of the monitoring and analytical schedule used in process kinetic evaluation of both the bench and pilot systems is presented in Table 2. This schedule was designed and utilized to permit the routine development of the following information:

1. Substrate balances (mass transfer rate basis).
2. Solids balances (mass transfer rate basis).
3. Flow balances.
4. Organic loading and removal rates.
5. Gas production rates and composition.
6. Fermentation control information (alkalinity, concentration, pH, and volatile acids concentration).

7. Nutrient balances (nitrogen and phosphorus).

The substrate, solids, and flow balances were developed on a weekly basis for each of the three units, and the balance data was reduced to develop a single set of experimental points per week, for each unit, for the kinetic characterizations. The solids and flow balances were used to characterize the performance of the solids separators. The nutrient balances were used to ascertain the sufficiency of influent nitrogen and phosphorus concentrations (in relation to the influent BOD concentration). The organic loading and removal rate data, and the gas production rate and composition data were used to develop overall measures of performance of the anaerobic contact units. The fermentation control information and the daily rate gas production rate data were used as real-time measures of the status of the fermentation and to establish the day-to-day operating changes required.

The data reduction process outlined above is illustrated by the example calculations presented in Appendix A of Section XII. The schedule of analyses presented in Table 2 was done using procedures of "Standard Methods"(16). The BOD and COD analyses were made on both unfiltered and filtered samples in the determination of total BOD and COD and soluble BOD and COD respectively.

Determination of Clarification Capacity

The objective of this phase of effort was to develop a relationship between clarifier performance and loading rate that could be used to examine tradeoffs between the recycle factor (C^o) attainable, and a loading parameter from which (1) the size and (2) the cost of the clarifier system could be established. The factors that were considered in the design of the experimental procedures were:

1. The inherent limitations of the pilot plant facility on the flexibility of operations.
2. The dependent (performance-related) and independent (loading-related) parameters to be selected.
3. The anticipated "shape" of the response surface relating the performance and loading parameters.

4. The selection of a suitable coagulant aid to improve the settleability of the mixed liquor solids, i.e., to improve the conversion of soluble, colloidal, and non-settleable suspended solids forms to a settleable form.

Evaluation of Experimental Variables - The pilot plant facility as utilized in the process kinetic characterization imposed several limitations on the clarification experiments, in that:

1. During the process kinetic characterizations, the maximum feed rate used (in the order of 190 liters/day or 50 gpd) resulted in maximal clarifier liquid loading rates of 0.204 cu m/day/sq m (5 gpd/sq ft), about 50 fold less than the maximal liquid loading rate capacity of 10.2 cu m/day/sq m (250 gpd/sq ft) desired for the clarification experiments.
2. Only a limited mass of suspended solids (approximately 50 kg of total suspended solids), at a constant concentration level of 12,000 to 14,000 mg/l was available for use in the clarification experiments; given the need to recycle (conserve) this inventory in the experiments, important considerations in evaluating the results of this effort were: (a) the recycle of coagulant aid with the recycled solids; (b) the buildup of a residual coagulant aid dose in the mixed liquor solids inventory; and (c) the potential adverse effects of a buildup to a coagulant-aid overdose situation.

The following steps were taken to resolve these limitations:

1. The pilot plant was modified to permit the recycle of both settled solids and clarified effluent back to the anaerobic fermenter - a measure necessary to conserve the essentially fixed supply of mixed liquor solids available.
2. Larger pumps were installed to permit the transfer of: (a) mixed liquor into the mix-head tank, and thence by gravity into the clarifier at a maximum surface loading rate of 10.2 cu m/day/sq m (250 gpd/sq ft); and (b) settled solids in the sludge recycle stream and clarified effluent back to the anaerobic fermenter.

The parameter selected as a measure of clarifier performance was the efficiency of total suspended solids (TSS) removal; when this efficiency parameter is expressed as a decimal fraction, the value of the recycle factor (C^o) is equal to 1-(TSS removal efficiency). The loading parameter selected was the solids loading rate (units of kg TSS/day/sq m or lbs/day/sq ft); this parameter can be related, through a scale-up factor, with the solids-handling-capacity parameter determined in laboratory zone settling assays. With TSS removal efficiency as the performance parameter, and solids loading rate as the independent parameter, the response surface relationship between these parameters can be defined as one of efficiency decreasing at an increasing rate with increasing loading rate, as documented in the literature(17).

The selection of a coagulant aid was done by a jar testing program that was designed with the objective of screening a wide range of high and low molecular weight polymers. Polymers as opposed to metal ion coagulants or lime were selected for the screening tests because of (1) the high fraction of dissolved and colloidal materials in the slops stream (as discussed in Section VI), and (2) the desire to minimize the dosing levels required and the associated waste sludge production. Polymers typically have a helical structure comprised of carbon chains with ionizing groups attached. These radicals ionize to effect a straightening of the chain by electrical repulsion. Colloidal solids can then interact with the straightened chain by electrical interaction and adsorption. As the charges neutralize, the polymer starts to coil again, and eventually forms a large enough floc to be susceptible to removal.

A basic problem with most polymers is that an overdose leads to an excess of charged sites being present in the aqueous solute, such that they cannot be neutralized(18). In this case, the straight chain structure is maintained, the system can become highly stable, and there is a diminished level of removal effected. Because of the potential overdosing problem associated with the use of polymers, and the anticipated accumulation of recycled residual polymer in the anaerobic contact process application, a primary criterion in the screening of candidate polymers was the identification of a polymer providing a high degree of conversion/separation over a wide range of polymer doses.

The polymer selected on the basis of jar testing was Nalco 627, a moderately cationic, high molecular weight polymer(19). With this polymer, it was possible to attain in excess of 80 percent TSS removals in the jar test beakers after 30 minutes of quiescent settling at specific polymer doses varying from 1 to 5 mg per gm TSS. At a mixed liquor TSS concentration of 10,000 mg/l, this specific dose range is equivalent to polymer doses of 10 to 50 mg/l.

Operating Procedures - The evaluation of clarification capacity was conducted as a two-step effort consisting of laboratory zone settling assays followed by pilot plant experimentation.

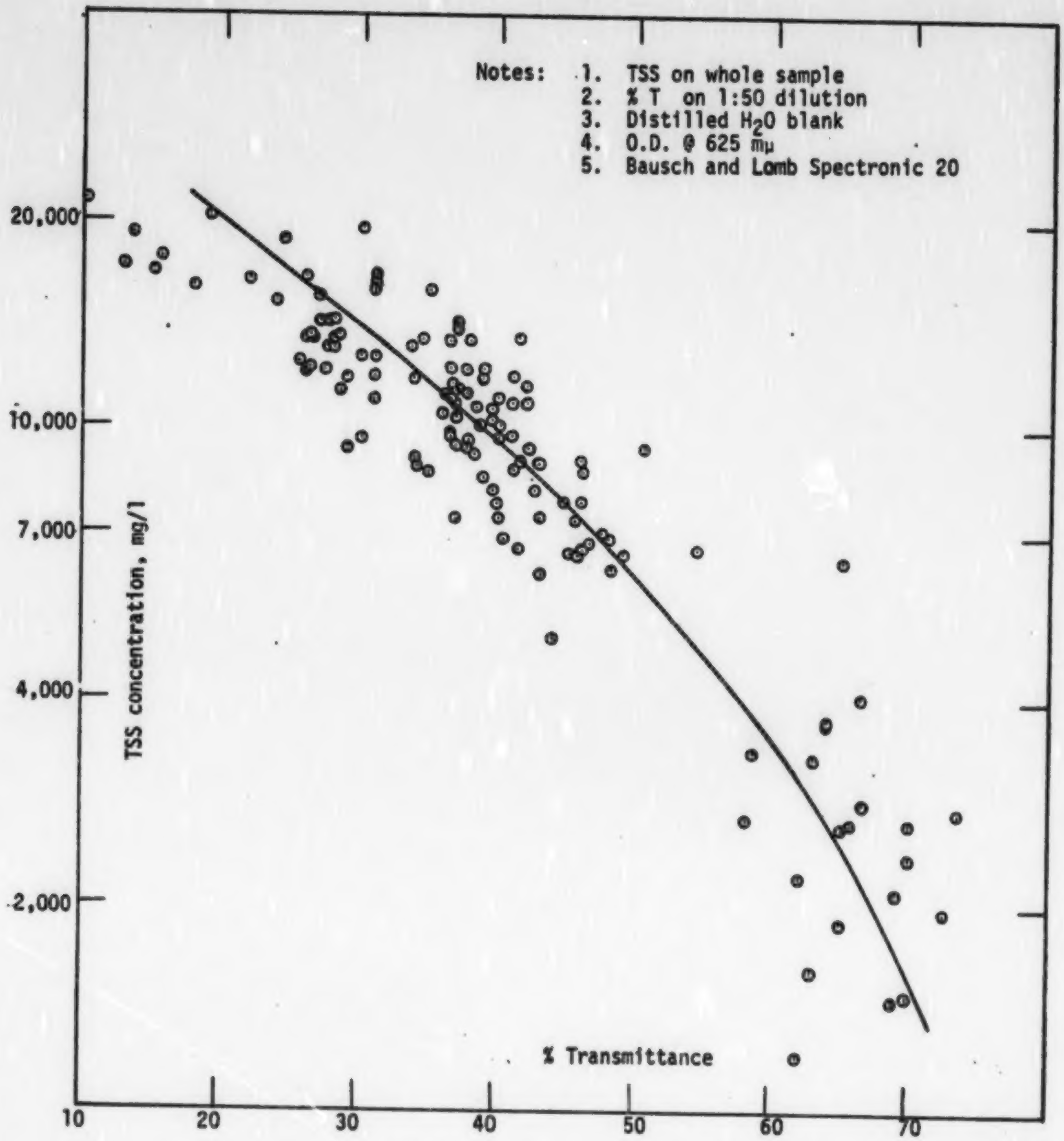
The zone settling assays were conducted with the objective of determining the solids handling capacity of the limiting layer as a function of specific polymer dose and initial TSS concentration in the suspensions. The zone settling assays were performed in one-liter graduate cylinders, using three mixed liquor slurries having initial TSS concentrations of approximately 6,000 mg/l, 12,000 mg/l, and 28,000 mg/l, respectively. Batches of slurries at each concentration were prepared by the dilution or settling of mixed liquor solids obtained from the pilot plant, and aliquots of each of the three slurries were pretreated at specific polymer doses of about 1 mg/gm TSS and 5 mg/gm TSS respectively. The observations made during each assay were interface height and elapsed time, from which settling curves were plotted. The settling curves were analyzed using the procedure described by Eckenfelder(17), for the determination of limiting layer solids handling capacity. The results of the zone settling assays are compared with the pilot plant experimental results (Section VI) as a basis for the selection of a design solids loading rate for the full-scale application.

Two consecutive series of experimental runs were conducted with the pilot plant to document the relationship between efficiency of TSS removal and solids loading rate. In the first series, no polymer addition was used, and runs were conducted at TSS loading rates of 48 kg/day/cu m (9.8 lb/day/sq ft) and 89 kg/day/cu m (18.1 lb/day/sq ft), equivalent to liquid loading rates of 2.64 cu m/day/sq m (65 gpd/sq ft) and 7.97 cu m/day/sq m (195 gpd/sq ft) respectively. In the second series, polymers were dosed throughout each of five runs conducted, at a specific polymer dose varying from 1 to 3 mg/gm, paced to the mass rate of transfer of mixed liquor TSS into the mix-head tank. The runs were conducted at TSS loading rates varying from 63.5 kg/day/sq m (13 lb/day/sq ft) to 125 kg/day/sq m (25.3 lbs/day/sq ft), equivalent to liquid loading rates varying from 4.87 cum/day/sq m (119 gpd/sq ft) to 8.48 cu m/day/sq m (297 gpd/sq ft). All of the above solids and liquid loading rates are expressed on an overflow basis.

The runs in each series were conducted in sequence over time periods varying from one to four hours per run. Observations were made at half-hourly intervals during the runs for the determination of the mixed liquor flow rate into the mix-head tank, clarified effluent flow rate, and sludge recycle flow rate. Samples were taken of each of the above streams at half-hourly intervals for determination of TSS concentrations. Aliquots of the samples taken from the mixed liquor stream entering the mix-head tank and from the clarifier effluent were analyzed immediately by an on-line optical density method (discussed below) for determination of estimated TSS concentrations in each of these streams on a real-time basis. The estimated TSS concentration in the mixer liquor stream was used to compute the mass rate of mixed liquid solids transfer into mix-head tank, as a basis for pacing the mass rate of polymer feed at the desired specific dose, and the estimated TSS concentrations in the clarified effluent and mixed liquor streams were used to calculate the TSS removal efficiency. All real time parameters developed on the basis of TSS concentrations determined with the on-line measurement technique were later recalculated using actual TSS measurements (which were available within 24 hours of sampling).

The on-line optical density method for estimation of TSS concentration was based on a correlation curve between (1) percent transmittance of a 1: 50 dilution of sample aliquot, as measured against a distilled water blank at a wave length of 625 mμ on a Bausch and Lomb Spectronic 20 instrument, and (2) the TSS concentration of the undiluted sample as determined by "Standard Methods"(16). The correlation curve used in the pilot plant experimentation is presented in Figure 7, and was developed by the determination of the optical density (according to the above procedure) and the TSS concentration of over 100 samples of mixed liquor and clarified effluent.

FIGURE 7. CORRELATION BETWEEN TOTAL SUSPENDED SOLIDS AND OPTICAL DENSITY



SECTION VI

RESULTS

CHARACTERISTICS OF SLOPS

Each of the slops batches used as a raw waste supply to the bench and pilot units was characterized as to organic, solids and nutrient composition, and several of the batches were analyzed for ionic composition. This effort was done to provide data on the influent loadings for the process performance analysis, and to permit a general characterization of this waste stream. From the data base developed, a summary of the wastewater parameters characterizing the slops is present in Table 3, and a summary of observations on the ionic composition of slops is presented in Table 4.

In the evaluation of these analytical results, it was found that the data sets on the soluble and total BOD and COD parameters (Table 3) can be categorized into two data distributions, each associated with one of the two distillation column modes in operation at the time the slops batches were collected. A similar differentiation (on the basis of distillation column operating mode) could not be developed with the data for the other parameters. The operating mode associated with the lower mean soluble and total COD and BOD concentrations was used about 75 percent of the time during the study, and the second mode was used the remainder of the time.

The mean soluble and total COD concentrations characteristic of slops obtained during the predominant operating mode were 72,000 mg/l and 74,800 mg/l, respectively (soluble COD: total COD ratio of 96 percent). The mean soluble and total COD concentrations characteristic of slops obtained during the other operating mode were 92,000 and 99,800 mg/l, respectively (soluble COD: total COD ratio of 0.92). Thus in the samples analyzed, both the soluble and total COD concentrations characterizing the slops during the latter mode of operation were about 30 percent greater, respectively, than those characterizing the slops in the predominant operating mode.

TABLE 3. SUMMARY OF SLOPS CHARACTERISTICS

Parameter	Mean	Range
Soluble COD (mg/l)	72,000[a]	67,100 - 75,700
	92,000[b]	81,100 -106,300
Total COD (mg/l)	74,800[a]	71,500 - 78,900
	99,800[c]	83,800 -115,500
Soluble BOD (mg/l)	26,500[b]	17,600 - 32,300
	47,400[d]	40,600 - 57,500
Total BOD (mg/l)	32,900[b]	19,800 - 41,900
	54,900[e]	45,800 - 67,000
Alkalinity (mg/l @ $CaCO_3$)[e]	912	806 - 1,320
Volatile Acids (mg/l @ HAc)[e]	4,920	3,610 - 5,920
pH[e]	4.36	4.28- 4.45
Solids (mg/l)[e]		
Total	83,500	70,200 - 95,800
. total fixed	20,500	19,400 - 22,200
. total volatile	63,000	50,700 - 73,600
Total dissolved	77,700	77,400 - 85,600
. fixed dissolved	19,800	17,900 - 21,500
. volatile dissolved	57,900	45,600 - 64,000
Total suspended	6,220	2,540 - 10,280
. fixed suspended	800	40 - 1,720
. volatile suspended	5,400	2,500 - 9,620
Nitrogen (mg/l @ N)[e]		
. total Kjehldahl	1,140	790 - 1,450
. organic	1,060	770 - 1,280
Total Orthophosphate (mg/l @PO_4)[d]	93	59 - 98

Notes: [a] Eight samples
[b] Six samples
[c] Seven samples
[d] Four samples
[e] Five samples

TABLE 4. IONIC COMPOSITION OF SLOPS
(Units of mg/l)

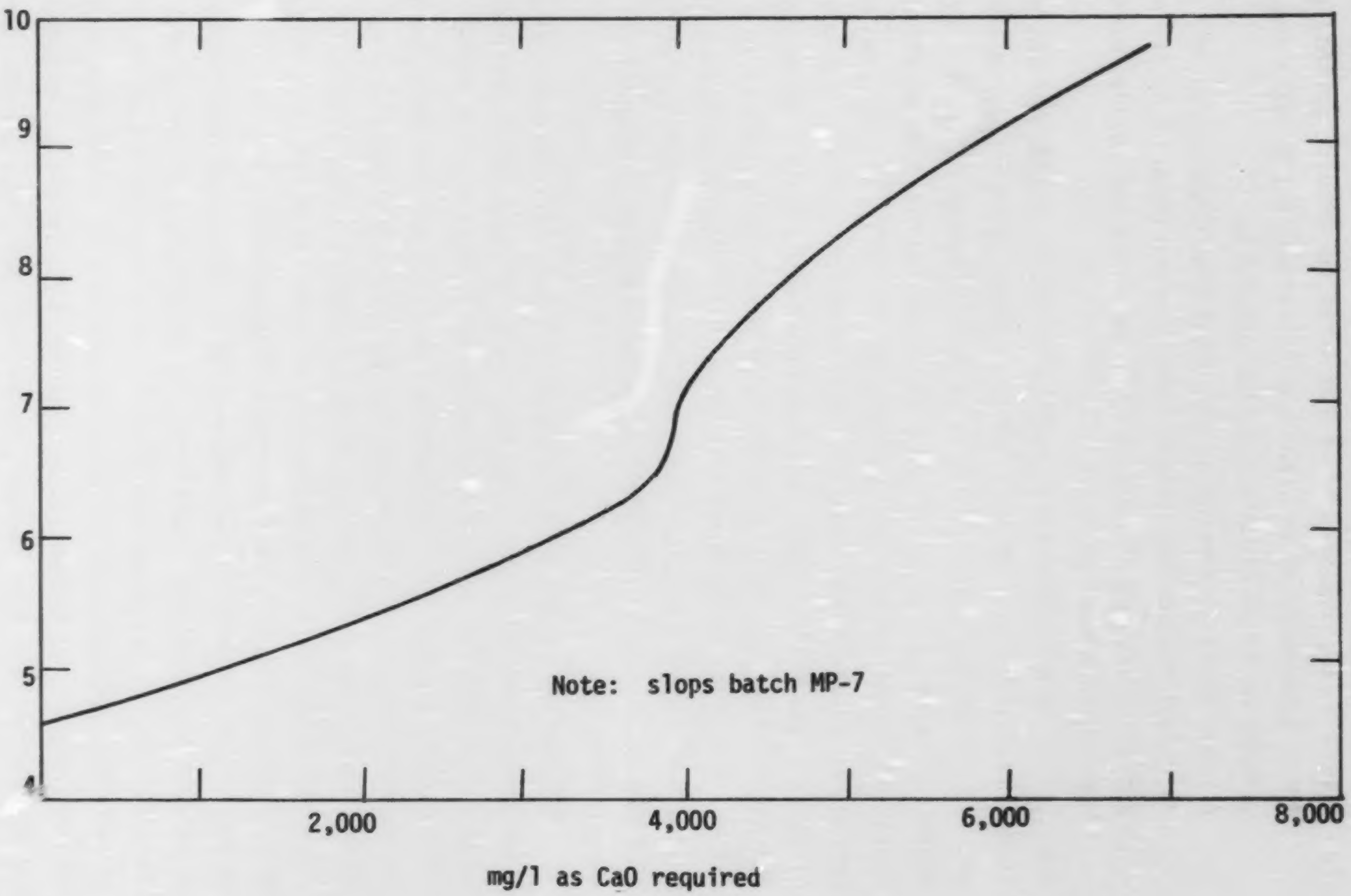
Constituent	Mean	Range	Observations
Zn	9.89	2.38 - 19.93	4
Cd	0.18	0.09 - 0.32	4
Pb	1.10	0.77 - 1.60	4
Fe	81.0	42.0 - 150.0	5
Na	372.	209. - 523.	5
Cu	32.8	2.0 - 124.	5
Co	0.60	0.19 - 0.76	4
Mn	10.6	2.38 - 15.6	4
Ca	2,088.	1,850. - 2,476.	4
Mg	824.	391. - 1,728.	5
Cr	0.30	0.25 - 0.33	4
K	4,259.	4,011. - 4,845.	5
Al	0.38	0.10 - 0.58	4
Cl	2,110.	1,330. - 4,400.	4
SO_4	4,120.	3,500. - 4,800.	3

The mean soluble and total BOD concentrations characteristic of slops obtained during the predominant operating mode were 26,500 mg/l and 32,900 mg/l respectively (soluble BOD: total BOD ratio of 0.80). During the second operating mode, the mean soluble and total BOD concentrations were 47,400 and 54,900 mg/l, respectively (soluble BOD: total BOD ratio of 0.86). Thus, both the soluble and total BOD concentrations characterizing the slops during the second mode of operation were about 20 percent greater than those in the first operating mode. Additionally, the average soluble and total BOD:COD ratios in slops generated during the first operating mode were about 0.35, whereas the same ratios for slops generated during the second operating mode were about 0.58.

The mean alkalinity of the slops samples analyzed was 912 mg/l $CaCO_3$, the mean volatile acids concentration was 4,920 mg/l as acetic acid, and the mean pH was 4.36 (Table 3). A titration curve, relating pH and lime dose required in the neutralization of a representative slops sample, is presented in Figure 8. The titration curve exemplifies, from an operational viewpoint, the interrelationship of the above parameters. The trend in the titration curve of Figure 8 is indicative of the existence of a buffering capacity in the slops at pH levels greater than 7.2. For this reason, the mixed liquor pH was maintained at or in excess of pH 7.2 throughout the experimental program by the addition of sodium bicarbonate as discussed in Section V. The lime dose required to neutralize the slops sample to pH 7.2 was 4,000 mg/l as CaO. At a lime dose of 4,000 mg/l as CaO, the calcium ion increment added to the slops (2,400 mg/l) represents an increase of approximately 115 percent in the average calcium content of the slops (2,088 mg/l) (Table 4).

The mean total solids concentration characteristic of the samples analyzed was 83,500 mg/l, approximately 25 percent of which was present as total fixed solids, and 75 percent of which was present as total volatile solids. Based on the mean values of concentrations reported for the dissolved and suspended solids parameters:

FIGURE 8. SLOPS NEUTRALIZATION RELATIONSHIP BETWEEN pH AND LIME DOSE



Note: slops batch MP-7

1. Approximately 7.5 percent of the total solids were present as total suspended solids (mean concentration, 6,220 mg/l) and 92.5 percent as total dissolved solids (77,700 mg/l).
2. Approximately 75 percent of the total dissolved solids were present as volatile dissolved solids (mean concentration of 57,900 mg/l) and the remaining 25 percent as fixed dissolved solids (mean concentration of 19,800 mg/l).

Thus, if it were assumed that the total volatile solids concentration is an implicit measure of the organic content (both biodegradable and refractory) of the slops stream, then approximately 85 percent of the organic content of the slops was present in soluble form as volatile dissolved solids (mean concentration, 57,900 mg/l) and the remaining 15 percent as volatile suspended solids, the latter consisting predominantly of spent yeast cells. These observations indicate that the organic content of the slops is predominantly of a soluble/colloidal nature rather than of a solids form, and suggest that the slops stream is well-suited for the anaerobic contact process application because of the high ratio of the concentrations of organic matter: suspended solids in the slops stream.

The mean total Kjehldahl nitrogen content characteristic of the slops samples analyzed was 1,140 mg/l, and the mean total orthophosphate concentration was 93 mg/l as phosphate, or 31 mg/l as phosphorus. From a biological treatability perspective, if it is assumed that the slops can be characterized as having an average total BOD concentration of 50,000 mg/l, then the "average" slops is characterized by a BOD:N:P ratio of 100:2.2:0.06. On comparison of this ratio with the rule-of-thumb ratio of 100:5:1 for a nutrient-balanced waste stream, it is apparent that there exists on the average a phosphorus deficiency of 94 percent and a nitrogen deficiency of 56 percent, in the stream. The effluents of the three experimental units were monitored routinely for nitrogen and phosphorus content to ensure that these nutrients were not limiting in the fermentations (Section V). It was found necessary to add phosphorus, but not nitrogen, to each of the slops batches throughout the process kinetic studies. Phosphorus was added in the form of a phosphate salt to satisfy a BOD:P ratio of 100:1.

The ionic composition of the slops is characterized by the analytical results summarized in Table 4. The ionic composition varied widely from sample to sample. The variation was attributable to factors such as changes in: (a) the composition of the feedstock molasses used in rum production, (b) the extent of chemical pretreatment of the molasses mixture prior to fermentation, and (c) the distillation operation. The mean iron concentration reported is 81 mg/l; this concentration comprises nearly two-thirds of the sum of mean concentrations of all heavy metal constituents in the slops analyzed. The mean sulfate concentration reported is 4,120 mg/l (equivalent to 1,373 mg/l as sulfide-sulfur). The ratio of mean heavy metal to mean sulfide sulfur concentration (approximately 9 percent) suggests the probability of precipitation of heavy metal sulfide compounds in the fermentation, the significance of which as a factor in controlling sulfide inhibition could not be evaluated in the scope of the present study. The mean potassium and calcium contents of the slops (4,260 mg/l and 2,100 mg/l respectively) are similar in magnitude to values reported by Bhaskaran(6) in the characterization of spent molasses washes from molasses distilleries in India.

PROCESS KINETICS

Data Base

The data base for the kinetic characterization was comprised of the weekly substrate, solids, and flow balances for the two bench-scale units and the pilot plant(20). The three biological systems were operated a total of 110 weeks, including acclimation operation over 16 week periods with each system. Exclusive of observations made during the startup phases in each of the three units, a total of 62 weekly sets of experimental data were available for use in the kinetic characterization. This data set was screened to eliminate data points not deemed representative of steady state operation. The criteria defined for steady state operation were as follows: less than 15 percent variation in flow rate; 10 percent variation in mixed liquor/effluent soluble COD concentration, 20 percent variation in mixed liquor/effluent

VSS concentration, and 20 percent variation in θ_T/C^o (sludge age), based on a comparison of a three-week running average of each measure with a given weekly value during the three-week period. A total of 14 sets of experimental data representative of steady-state were identified using these criteria, seven from operation of the two bench units, and seven from the pilot plant operations.

The characteristics of the fermentations in these units at each steady state point are summarized in Tables 5 and 6, and illustrated by the data presented in Figure 9. The values of θ_T/C^o (sludge age) at which steady state was obtained varied from 35 to 220 days, and the values of θ_T (hydraulic residence time) varied from 19 to 221 days.

The soluble COD loading rate, expressed in rational units of mass/time-mass of MLVSS, varied from a minimum of about 0.26 kg/day/kg MLVSS (0.26 lb/day/lb MLVSS) at a θ_T/C^o of 220 days, to a maximum of 0.74 kg/day/kg MLVSS at a θ_T/C^o of 35 days. The soluble BOD loading rates corresponding to the above were 0.059 kg/day/kg MLVSS at θ_T/C^o of 220 days, and 0.45 kg/day/kg MLVSS at θ_T/C^o of 35 days. Thus, the maximum soluble COD loading rate observed under steady state conditions (0.74 kg/day/kg MLVSS) was approximately 25 percent less than the maximal loading rate of one kg/day/kg MLVSS recommended for startup (Section V).

On a volumetric basis, the soluble COD loading parameter varied from a minimum of about 0.4 kg/day/cu m (0.024 lb/day/cu ft) at θ_T/C^o of 220 days, to a maximum of 3.6 kg/day/cu m (0.23 lb/day/cu ft) at θ_T/C^o of 53 days, as shown in Figure 9. The volumetric soluble BOD loading rates varied from 0.086 to 1.19 kg/day/cu m(0.005 to 0.074 lb/day/cu ft), increasing with decreasing θ_T/C^o values in a pattern similar to that for the soluble COD pattern.

The alkalinity of the mixed liquors varied from 8,500 to 11,000 mg/l as $CaCO_3$, and the pH levels varied from 7.11 to 7.32. The ranges of values of each of these parameters reflected the effect of buffer addition as necessary to maintain the pH of the mixed liquors at about 7.2 (Section V). The volatile acids concentrations in the fermentations

TABLE 5. STEADY-STATE OPERATIONS: SUMMARY OF LOADING RATE, MIXED LIQUOR, AND EFFLUENT CHARACTERISTICS

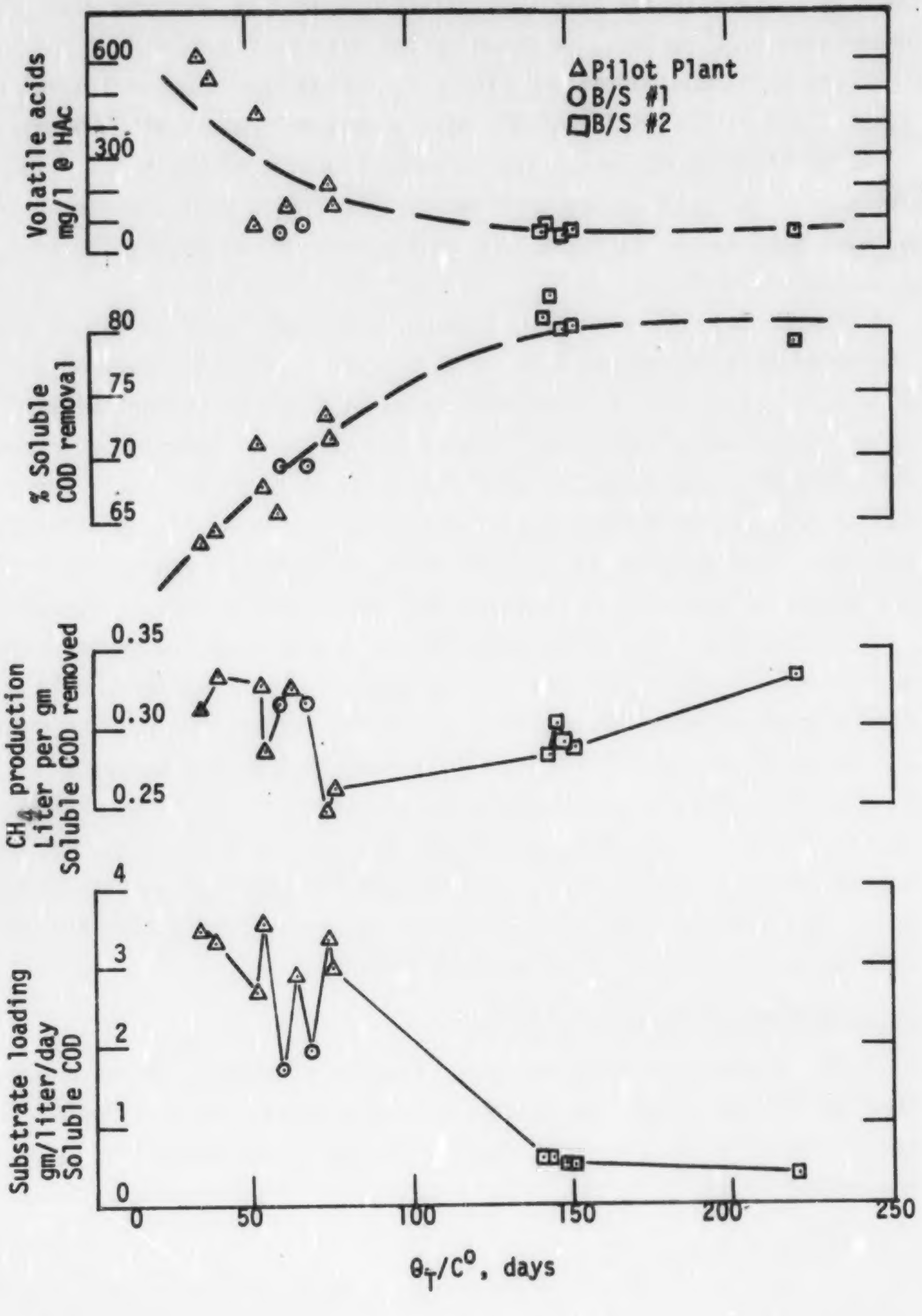
			Loading Rates				Mixed Liquor Characteristics						
Syst.	Wk.	θ_T (days)	Kg Sol COD / Day-Cu m	Kg Sol BOD / Day-Cu m	Kg Sol COD / Day-KgMLVSS	Kg Sol BOD / Day-KgMLVSS	Sol COD (X_1^N) mg/1	VSS (X_1^0) mg/1	MLTSS mg/1	pH[1]	Alkalinity (mg/1 $CaCO_3$)	Vol. Acid (mg/1 HAc	Effluent Sol COD (X_2^N) mg/1
B/S 1	25	35.1	1.69	1.03	0.729	0.443	21,230	2,320	3,287	7.21	9,170	94	20,070
	26	35.4	1.95	1.19	0.738	0.449	21,250	2,647	3,867	7.20	9,410	87	19,390
B/S 2	29	221	0.38	0.086	0.261	0.059	17,790	1,458	2,120	7.32	9,280	59	--
	30	149	0.56	0.127	0.380	0.086	16,460	1,475	2,400	7.31	9,180	60	20,910
	31	147	0.57	0.129	0.390	0.088	16,790	1,462	2,200	7.23	9,080	57	16,030
	32	143	0.58	0.132	0.399	0.089	15,110	1,480	2,713	7.24	8,880	75	16,300
	33	145	0.58	0.131	0.395	0.089	14,500	1,469	2,518	7.27	8,570	54	--
PP	21	25.2	2.67	0.701	0.622	0.163	18,950	4,293	6,587	7.15	8,710	92	16,530
	22	20.6	2.90	0.760	0.557	0.146	22,770	5,200	7,360	7.12	8,780	143	21,780
	24	19.0	3.57	0.935	0.700	0.183	21,570	5,100	6,990	7.16	8,900	436	20,430
	25	20.3	3.33	0.872	0.582	0.152	23,810	5,720	7,853	7.16	8,860	549	23,430
	26	19.3	3.44	0.901	0.686	0.159	24,420	5,020	7,333	7.11	9,070	618	22,900
	33	31.0	3.36	0.991	0.560	0.165	26,260	6,000	8,600	7.20	11,010	220	28,040
	34	36.9	3.01	0.888	0.497	0.147	29,450	6,067	9,027	7.19	10,820	154	28,270

Note: [1] pH controlled by buffer addition

TABLE 6. STEADY STATE OPERATIONS: SUMMARY OF OPERATING AND PERFORMANCE CHARACTERISTICS

			Gas Production Parameters					Sol. COD Removal		
System	Week	θ_T (days)	Recyc. Factor (C^0)	Daily gas Production (1/day)	% CH_4	1 CH_4 per gm VSS-day	1 CH_4 per gm Sol COD removd	Influent Mixed Liquor(%)	Influent Separator Effluent (%)	$\frac{\theta_T}{C^0}$ (days)
B/S 1	25	35.1	0.604	14.7	53.7	0.160	0.317	69.3	71.0	58.1
	26	35.4	0.523	16.7	54.6	0.106	0.316	69.3	72.0	67.7
B/S 2	29	221	1.004	16.6	54.4	0.291	0.333	78.8	--	220.
	30	149	0.992	18.0	58.7	0.336	0.289	80.4	75.0	150.
	31	147	1.000	17.5	59.9	0.337	0.290	79.9	80.9	147.
	32	143	1.012	18.6	58.5	0.345	0.282	82.0	80.5	142.
	33	145	1.002	19.8	58.8	0.372	0.303	82.7	--	144.
PP	21	25.2	0.410	2,053.	58.3	0.147	0.329	71.7	75.3	51.5
	22	20.6	0.372	2,045	56.9	0.120	0.324	66.0	67.5	63.0
	24	19.0	0.359	2,386	54.8	0.131	0.285	67.8	69.5	53.0
	25	20.3	0.525	2,379	59.4	0.131	0.333	64.5	65.1	38.7
	26	19.3	0.598	2,181	59.5	0.136	0.313	63.6	65.9	34.6
	33	31.0	0.417	2,055	58.0	0.105	0.249	74.7	73.0	74.3
	34	36.9	0.492	1,860	58.1	0.094	0.265	71.6	72.7	75.0

FIGURE 9. STEADY-STATE CHARACTERISTICS-RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS



varied from less than 100 mg/l (as acetic acid) at θ_T/C^o values in excess of 100 days, to maximal values of 600 mg/l at θ_T/C^o values of 35 days. The volatile acid concentration tended to increase with decreasing θ_T/C^o values, as shown by the data in Figure 9.

The MLVSS concentrations (Table 5) varied from a minimal level of about 1,500 mg/l at θ_T/C^o of 220 days to maximal levels of 6,000 mg/l at θ_T/C^o of 35 to 40 days. The volatile fraction of the MLTSS, i.e., ratio of MLVSS:MLTSS concentrations for the steady state points varied between 0.55 and 0.73; for all 14 data points the average volatile fraction was 0.67.

The soluble COD removal efficiency (influent → mixed liquor basis) increased with increasing θ_T/C^o from a minimal value of about 65 percent at a θ_T/C^o of 35 days to a maximal level of about 80 percent at θ_T/C^o values in excess of 150 days. From a comparison of soluble COD removal efficiencies computed on (1) an influent → mixed liquor basis and (2) an influent → separator effluent basis (Table 5), it is evident that the additional soluble COD removal obtained in the separator sector is insignificant relative to removals obtained in the anaerobic fermenters.

The methane (CH_4) content of the gas produced in the fermentations varied from 53 to 60 percent, and averaged 57.4 percent for the 14 steady state points. The methane production varied from 0.28 to 0.33 liters of CH_4 per gm of soluble COD removed (4.5 to 5.3 cu ft/lb soluble COD removed), and averaged 0.302 liter/gm soluble COD removed (4.8 cu ft/lb soluble COD removed) for all observations. That is, an average of 86.3 percent of the soluble COD (or about 82 percent of the total COD) removed from the feed stream was recovered in the form of methane during the steady state observations.

Determination of Process Kinetics

For purposes of determining the kinetic constants, it was assumed that the VSS parameter represented viable biomass concentration (X_1^o), and the soluble COD parameter represented the rate-controlling substrate concentration (X_1^N).

Equations 13 and 14 were used to analyze the steady-state data for determination of the kinetic constants. Presented in Figure 10 is a plot of Equation 13, in which the intercept is the inverse of the value of Y^o, and the slope represents the ratio K^D/Y^o. The steady-state data are presented in Figure 11 for determination of the values of K and k^m using Equation 14. The intercept of the straight line fit in Figure 11 is the inverse of the value of k^m, and the slope is equal to K/k^m.

The values of the kinetic constants obtained from this analysis are as follows (the calculations required are presented in Appendix A of Section XII):

Y^o = 0.225 mg VSS/mg soluble COD
K^D = 0.0667 days^{-1}
K = 12,270 mg/l soluble COD
k^m = 0.129 days^{-1}

In order to provide some basis of comparison for the values of the kinetic constants, a summary of the kinetic constants determined in this and prior studies is presented in Table 7. Included are data for a packinghouse waste(11,12), synthetic milk waste(9), and the results reported by Andrews and Pearson(21).

The value of k^m for the slops and synthetic milk wastes are similar in magnitude (0.13 days^{-1}). Given the 10°C difference in temperature used in these two studies, it was anticipated based on observations by Schroepfer and Ziemke(11,12) that the anaerobic stabilization rate for the slops would have been about double that for the synthetic milk waste. However, since the temperatures used by Andrews and Pearson(21), Schroepfer et al(11), Schroepfer and Ziemke(12), and in this study were nearly equal, it would appear that k^m is influenced by substrate composition as well as temperature.

The K^D value obtained for slops treatment by the anaerobic contact process is nearly equivalent to that reported for the synthetic milk waste(9). The Y^o value obtained for rum slops is greater than that reported by Andrews(21) for the methane organism dominated system and less than that reported by Gates et al(9) for the synthetic milk waste. The variation of Y^o values appears to be attributable both to

FIGURE 10. RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS-DETERMINATION OF VALUES OF Y^O AND K^D FOR STEADY-STATE DATA

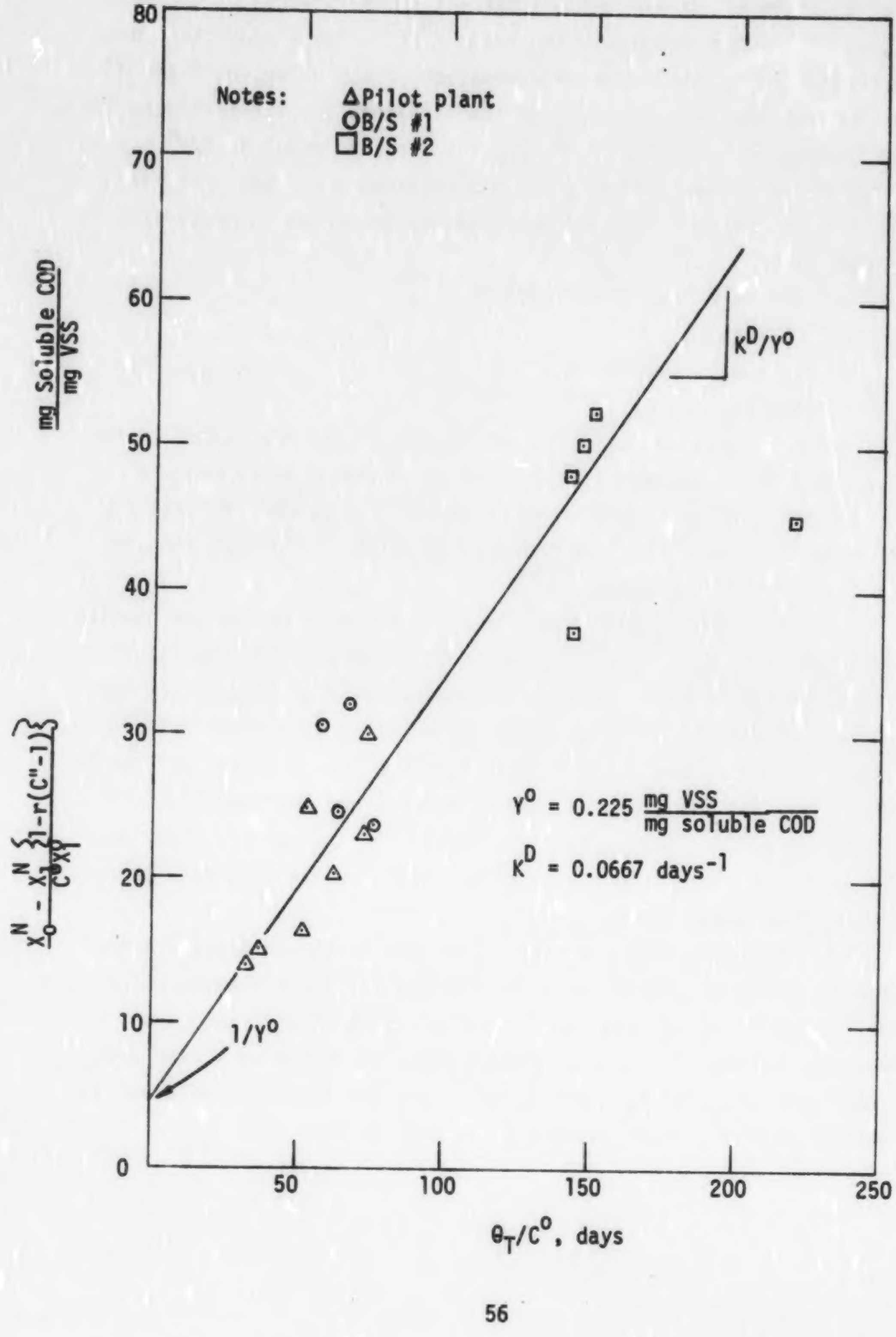


FIGURE 11. RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS DETERMINATION OF VALUES OF k^m AND K FOR STEADY-STATE DATA

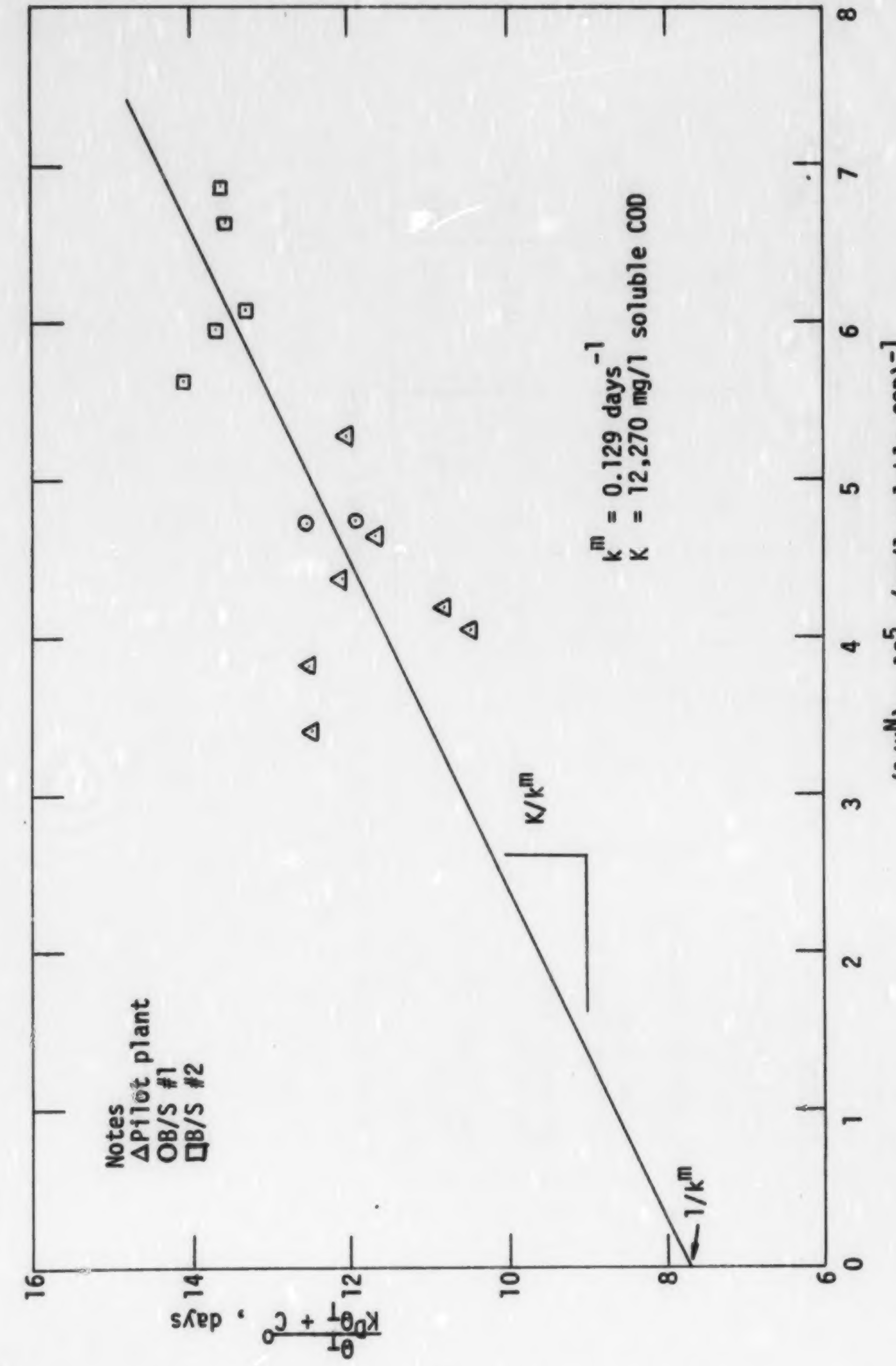


TABLE 7. VALUES OF THE KINETIC CONSTANTS (k^m, K^D, K, and Y^o) IN VARIOUS STUDIES

Source	Constants				
	k^m(days^{-1})	K^D(days^{-1})	K(mg COD/1)	$Y^o\left(\frac{\text{mg VSS}}{\text{mg COD}}\right)$	Temp (°C)
Packinghouse waste (11,12)	0.24	0.17	5.5	0.76	35
Anaerobic digester (21)					
Acid-formers	1.33	0.87	--	0.54	38
Methane formers	1.33*	0.02	--	0.14	38
Synthetic Milk Waste (9)	0.14	0.07	24.3	0.37	20-25
Rum Slops (this study)	0.129	0.0667	12,270	0.225	33-36

*At least some of the methane formers exhibited this value temperature and to different relative combinations of organisms existing in the different systems. The K value for the rum slops is several hundredfold greater than the value reported for the other wastes. That K did change with temperature and/or substrate composition is apparent. However, too little is known about the effect of transport-controlled mechanisms to support any definite conclusions about the impact of the environment on K.

Operational Relationship

Equation 15 and the values of the constants for rum slops treatment by the anaerobic contact process were used to formulate the operational relationship defining process performance as a function of θ_T and C^o, as presented in Figure 12. The measure of process performance is the soluble COD concentration of the process effluent, X_1^N. The operational relationship is a hyperbola which becomes asymptotic to an ordinate value equal to $K^D K/(k^m - K^D)$ and to an abscissa value equal to $1/(k^m - K^D)$. The equation describing the operational relationship is as follows:

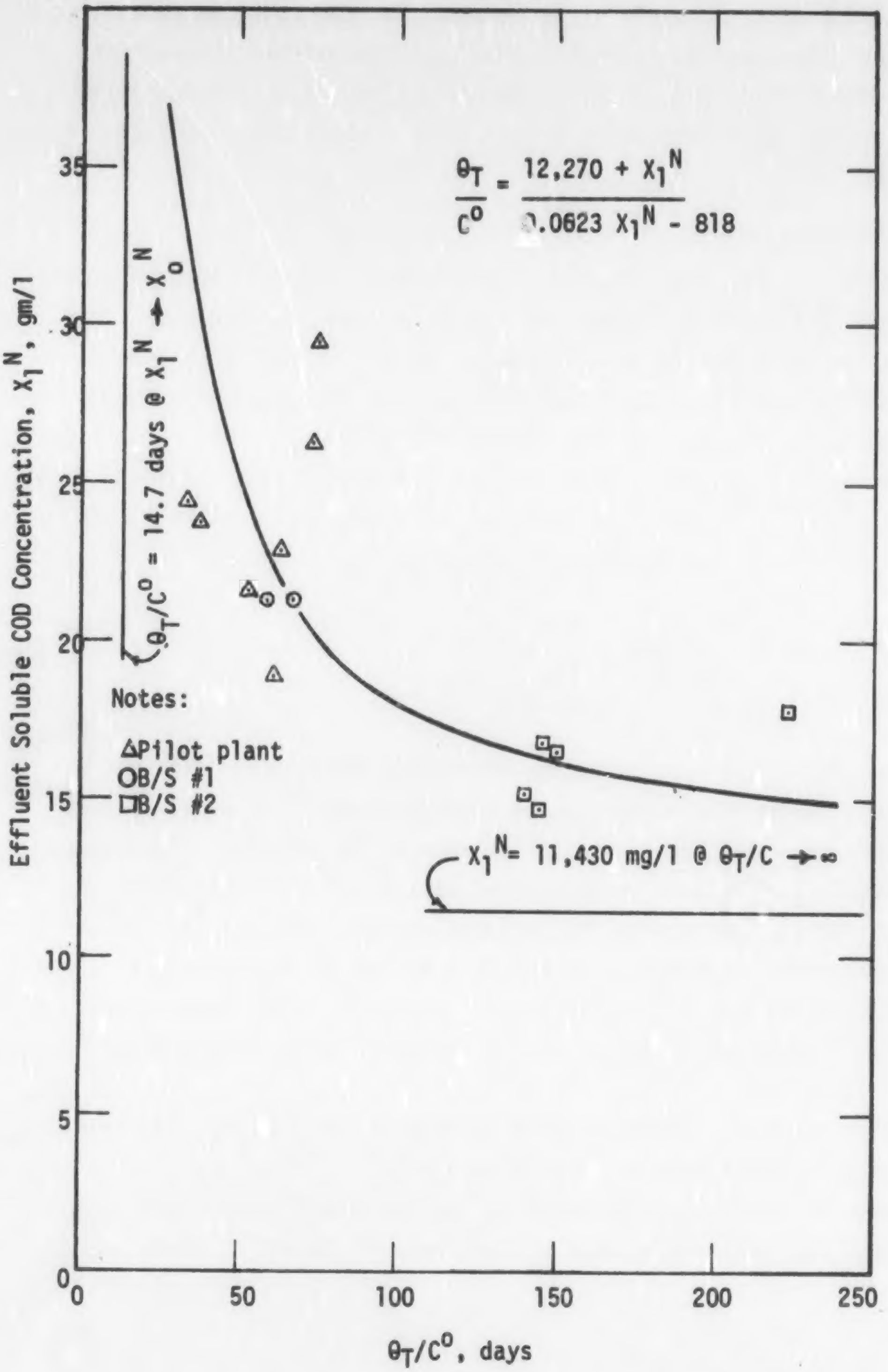
$$\frac{\theta_T}{C^o} = \frac{12{,}270 + X_1^N}{0.0623\ X_1^N - 818} \qquad (16)$$

The value of the ordinate asymptote represents the minimum effluent substrate concentration which can be obtained when C^o is equal to zero (i.e., 100 percent biomass recycle) or when θ_T is infinity. The value of the abscissa asymptote is approached as the effluent substrate concentration X_1^N approaches the influent concentration (X_o^N), i.e., when biomass washout occurs. In the case of rum distillery slops, these asymptotic values are X_1^N = 11,430 mg/l and θ_T/C^o = 14.7 days, respectively.

Three regions of operation can be defined with the operational curve (Figure 12):

1. A zone of stable operation (small ratio of $\Delta X_1^N : \Delta\theta_T/C^o$), existing at θ_T/C^o values greater than 75 days.
2. A zone of transition (in which X_1^N varies significantly with changes in θ_T/C^o), existing at θ_T/C^o values between 40 and 75 days.

FIGURE 12. UNIFIED OPERATIONAL RELATIONSHIP FOR RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS



3. A zone of instability (occurring at θ_T/C^o) values less than 40 days), in which the effluent quality rapidly deteriorates.

Within the zone of stable operation, effluent soluble COD concentrations of less than 20,000 mg/l were obtained. This effluent concentration is equivalent (for a slops feed stream concentration of 80,000 mg/l soluble COD), to a soluble COD removal efficiency of 75 percent.

CLARIFICATION CAPACITY

Both laboratory zone settling tests and pilot clarification tests were conducted according to procedures described in Section V, to permit the development of the desired relationship between clarifier performance and loading.

Zone Settling Tests

The results of the six zone settling assays, conducted using three initial MLTSS concentrations and two polymer doses per initial MLTSS concentration, are presented in Table 8. The actual MLTSS concentrations at which the tests were conducted were approximately 6,500 mg/l, 12,000 mg/l, and 28,000 mg/l, the lower and upper values being representative of the anticipated range of MLTSS concentrations to be carried in a full scale application. The data on solids loading rate in the limiting layer (Table 8) were calculated using the zone settling curves (plots of interface height vs settling time) for each test, and an assumed underflow concentration of 50,000 mg/l (TSS). The allowable liquid loading rate was then calculated from the initial MLTSS concentration and solids handling rate data for each test. The allowable liquid loading rate and initial MLTSS data have been plotted in Figure 13 to determine, for the laboratory data, a relationship between maximum rate of clarification as related to the settling velocity of the limiting layer, and initial MLTSS concentration.

A trend can be defined between allowable liquid loading rate and initial MLTSS concentration at the lower range of polymer doses (0.63 to 1.44 mg/gm MLTSS), as exemplified by the best-fit curve in Figure 13. The allowable liquid loading rate decreased from 26 cu m/day/sq m (635 gpd/sq ft) at an MLTSS concentration of 7,000 mg/l to 11.5 cu m/day/sq m (260 gpd/sq ft) at 30,000 mg/l.

TABLE 8. SUMMARY OF ZONE SETTLING DATA

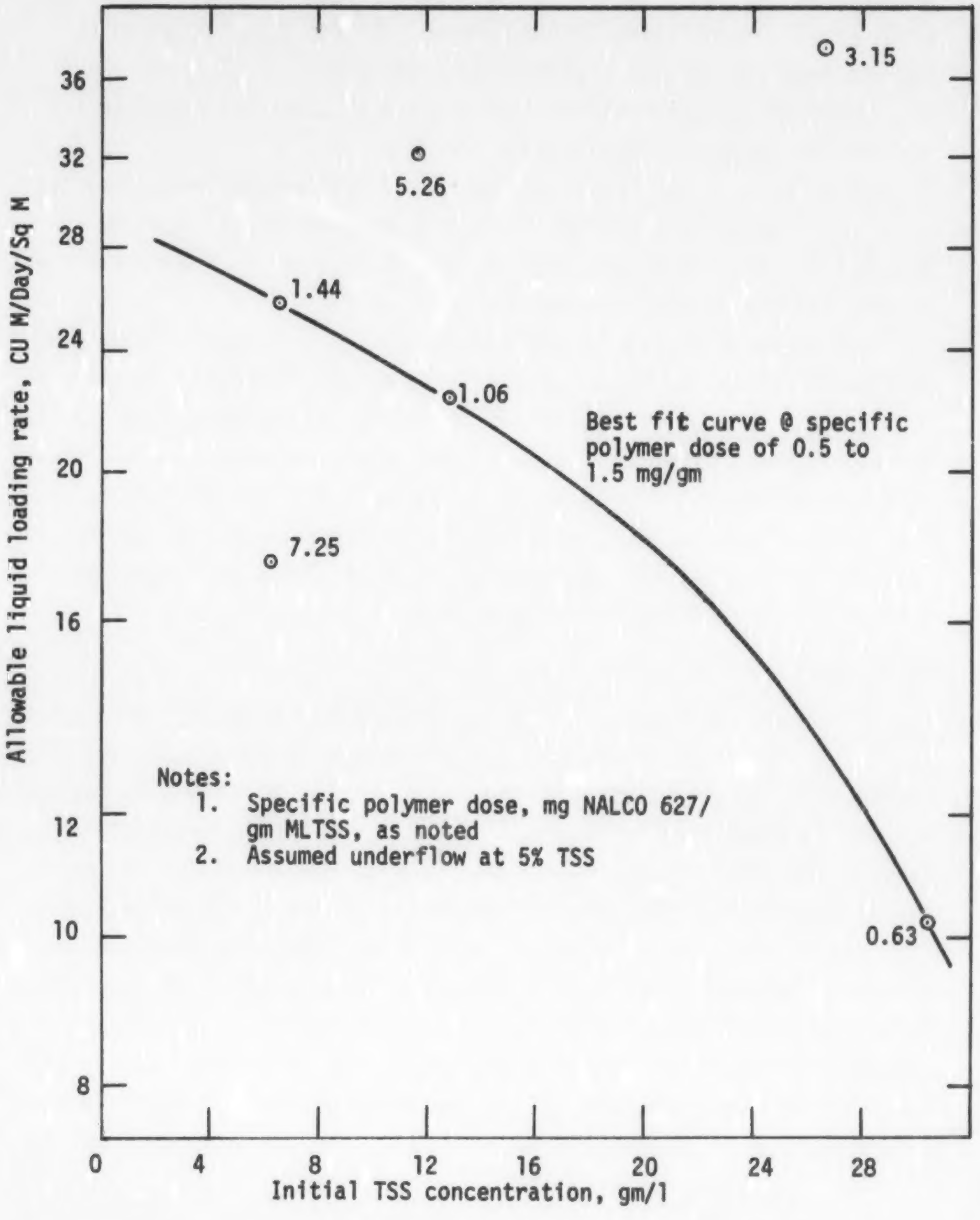
Test #	Initial MLTSS (mg/1)	Poly-mer dose[1] (mg/gm)	Solids Loading in Limit. Layer[2] (kg/day-sq m)	Allowable Liquid Loading Rate (cu m/day/sq m)	Superna-tant TSS[3] (mg/1)	% TSS Removal
1	6,770	1.44	178	26.4	1,450	78.5
2	6,295	7.25	112	17.9	3,130	50.2
3	12,850	1.06	293	22.8	1,200	90.7
4	11,635	5.26	381	32.8	1,700	85.4
5	30,527	0.63	317	10.5	2,230	92.7
6	26,594	3.15	1,001	37.7	2,930	89.0

Notes: [1] Nalco 627

[2] Assumed underflow TSS concentration - 50,000 mg/1

[3] Sampled after 30 minutes of settling

FIGURE 13. ALLOWABLE MIXED LIQUOR SURFACE LOADING RATE VS MLTSS CONCENTRATION - ZONE SETTLING



Insufficient points were available to define similar trends for liquid loading rates associated with the high polymer doses. However, it is apparent from the data presented in Figure 13 that at polymer doses greater than 1.5 mg/gm MLTSS, the allowable liquid loading rate at any given MLTSS concentration (in comparison with the trend obtained at polymer doses in the range of 0.63 to 1.44 mg/gm):

1. Increased with increasing polymer dose to a maximal level at a polymer dose of 3 mg/gm, and;
2. From this maximal level then decreased with polymer doses greater than 3 mg/gm to a minimal level at a polymer dose of 7.25 mg/gm.

This optimal dose effect was anticipated in view of prior experience with polymer applications as discussed in Section V.

The decrease in polymer effectiveness with increased polymer dose is also evidenced by the TSS removal efficiency data presented in Table 8. That is, the percent removal of initial MLTSS, as determined by sampling the supernatants after 30 minutes of settling, decreased from 85 percent at a polymer dose of 5.26 mg/gm to 50 percent at a dose of 7.25 mg/gm. Thus, the results of the zone settling assays confirmed the conclusions from the jar testing that the polymer selected (Nalco 527) would be effective at doses from 1 to 5 mg/gm MLTSS.

Pilot Plant Clarification Tests

The pilot plant clarification tests were conducted in two series of runs. The first series was done without polymer addition, and the second series with polymer addition, using the procedures described in Section V. A summary of the results obtained in each series/run is presented in Table 9, and the relationships between solids removal efficiency and solids loading rate derived from these results are presented in Figure 14.

The information presented in Table 9 includes: run duration; average cumulative polymer dose; liquid and solids loading rates; and average MLTSS removal efficiency over the duration of the run. The average cumulative polymer dose reported in Table 9 for each Series II run is equal to the cumulative mass of polymer added per mass of MLTSS in the entire pilot plant system, during each run. The actual rate of polymer

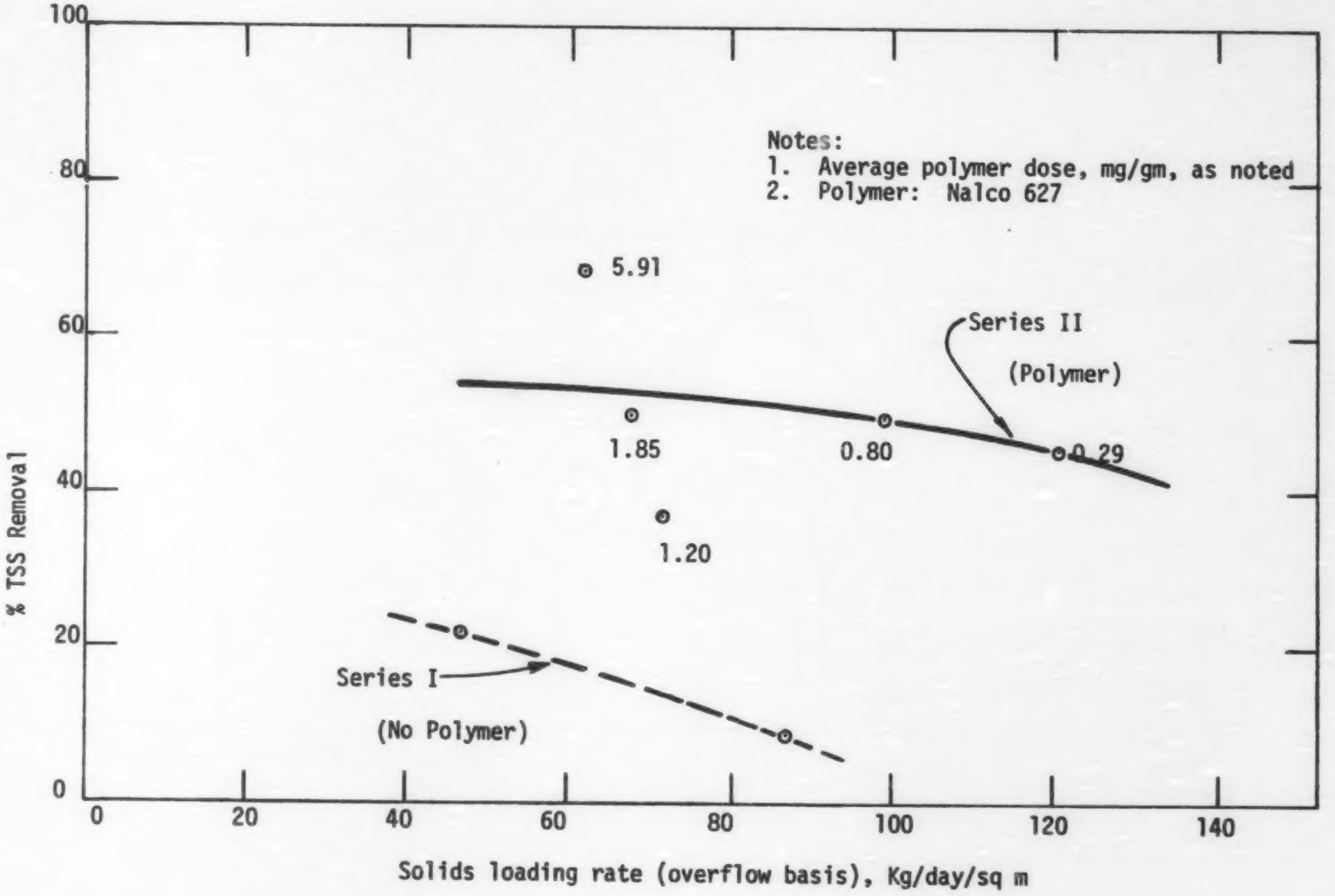
TABLE 9. SUMMARY OF RESULTS-PILOT PLANT CLARIFICATION TESTS

Series run	Duration (hours)	Average Cumulative Polymer Dose[a]	Liq. Load. Rates (cu m/da/sq m)		Sol Load Rate (kgMLTSS/da/sq m)		Influent MLTSS (mg/l)	Average MLTSS Removal Efficiency (%)
			Overflow basis	Total basis[b]	Overflow basis	Total basis[b]		
I-1	1.0	--	2.63	5.48	48.0	100.0	10,833	21.8
I-2	4.0	--	7.96	10.8	88.5	120.3	11,144	8.8
II-1	2.0	0.29	8.45	11.3	124.	165.	14,660	45.3
2	3.0	0.80	7.65	10.5	102.	140.	13,330	49.8
3	1.5	1.20	5.08	7.94	73.5	115.	14,500	36.7
4	4.0	1.85	5.03	7.88	69.6	109.	13,880	50.1
5	1.5	5.91	4.86	7.71	63.7	101.	13,150	68.6

Notes: [a] Cumulative mass of polymer (Nalco 627) added per mass of MLTSS in pilot plant

[b] Based on overflow as clarified effluent plus sludge recycle flow

FIGURE 14. RELATIONSHIP BETWEEN SOLIDS REMOVAL EFFICIENCY AND SOLIDS LOADING RATE (WITH AND WITHOUT POLYMER)



dosing into the mixed liquor stream during the Series II runs was set at from 1 to 3 mg/gm of MLTSS transferred into the mix-head tank. The "overflow" liquid and solids loading rate data reported in Table 9 were calculated on the basis of the clarified effluent flow rate, and the "total" liquid and solids loading rate data were calculated using the sum of the clarified effluent and sludge recycle flow rates. The influent MLTSS concentrations averaged about 11,000 mg/l during the Series I runs and 14,000 mg/l during the Series II runs. The solids loading rates for each run were computed using the corresponding liquid loading rates and influent MLTSS concentrations as reported in Table 9.

The effectiveness of the polymer addition in improving the settleability characteristics of the mixed liquor solids is evident from the relationships presented in Figure 14 for the Series I and II data. In the Series I relationship, the TSS removal efficiency decreased from 22 percent at a solids loading rate (overflow basis) of 48 kg/day/sq m (9.8 lbs/day/sq ft) to a level of eight percent at a solids loading rate of 88.5 kg/day/sq m (18 lbs/day/sq ft); i.e., without polymer addition, the mixed liquor solids were effectively unsettleable at these solids loading rates. With polymer addition in the Series II runs, the TSS removal varied from an average of 52 percent at a solids loading rate of about 70 kg/day/sq m (14.3 lb/day/sq ft) to 45 percent at 120 kg/day/sq m (24.6 lb/day/sq ft). The solids loading rates of 70 and 120 kg/day/sq m were equivalent to liquid loading rates (overflow basis) of 5 cu m/day/sq m (123 gpd/sq ft) and 8.4 cu m/day/sq m (205 gpd/sq ft) respectively. Additionally, at the solids loading rate of 70 kg/day/sq m, the TSS removal efficiency tended to increase with increasing cumulative polymer dose in the pilot plant system, varying from 37 percent at a cumulative dose of 1.2 mg/gm to nearly 70 percent at a cumulative dose of 5.9 mg/gm.

Based on the preceding observations, it is evident that polymer addition is a requisite for any full-scale application, and that the "residual" effect of the polymer recycled in the system (as measured in the present study by the average cumulative dose), was beneficial, within the range of average cumulative doses at which the tests were made, to the settleability of the mixed liquor solids.

A major equipment limitation in the pilot plant, which could not be resolved in the course of the pilot plant testing, was the clogging of several of the funnel weirs in the clarifier (Figure 6). Because of the clogging, only 0.32 m (1.05 ft) of the design weir length of 2.57 m was usable, and the operative weir loading factor was equal to 3.13 cu m/day/meter of weir length per cu m/day/sq m of liquid loading rate (250 gpd/ft per cu m/day/sq m). Thus, during the Series II runs, the weir loading rates varied from approximately 15 cu m/day/m (1,210 gpd/ft) to 26.4 cu m/day/m (2,120 gpd/ft). While these weir loading rates are low in comparison with rates used in design of primary sedimentation tanks (greater than 50,000 gpd/ft), the malfunction of the pilot plant weirs disrupted the flow distribution pattern at the surface of the clarifier. Because of the short circuiting of the surface flow pattern in the clarifier, it is anticipated that TSS removal efficiencies on a full-scale application can be increased at least 20 percent by the careful design of the flow pattern in the clarification units, and the use of weir loading rates of less than 12.4 cu m/day/m (1,000 gpd/ft).

A comparison can be made between the zone settling results and the pilot plant clarification tests by using the curve of Figure 13, to determine the allowable solids loading rate as the product of the allowable liquid loading rate and the initial TSS concentration. The allowable solids loading rate in the initial TSS concentration range of 10,000 to 14,000 mg/l, as determined from the zone settling results, is 239 to 302 kg/day/cu m (49 to 62 lbs/day/sq ft). Allowing a scale-up factor of 3 to 4 to conservatively account for scale-up differences (hydrodynamics) between the laboratory apparatus and the pilot plant, then a comparability of clarification efficiency at the pilot scale would be expected at a pilot plant solids loading rate (overflow basis) of 80 to 100 kg/day/cu m (16.4 to 20.5 lbs/day/sq ft). Because of the stability of pilot plant performance observed over this range of pilot plant solids loading rates (Figure 14), and the above cited weir loading rates in the pilot plant clarifier, the comparability of the zone settling results and the pilot plant results was reasonable. On this basis, it was concluded that, with a properly designed full-scale clarification unit, a clarification efficiency of 70 percent can be attained at a solids loading rate of 100 kg/day/sq m (20.5 lbs/day/sq ft) and a weir loading rate of 12.4 cu m/day/m (1,000 gpd/ft) or less, with polymer doses in the range of 1 to 5 mg/gm MLTSS in the clarifier influent stream.

SECTION VII

DESIGN AND OPERATIONAL ANALYSIS

The results of the bench and pilot studies provide the process documentation and experience for examining the design, operational, and economic aspects of the full-scale application of the anaerobic contact process for slops treatment. An analysis of the design and operational requirements for the application is presented in this section, and an economic analysis of the required system is presented in Section VIII.

The three elements of the design and operational analysis presented herein are:

1. The selection and specification of a steady-state operating region for the application;
2. The specification of operating requirements;
3. The specification of a process flow sheet and design criteria for the physical system.

STEADY-STATE OPERATING REGION

The specification of a steady-state operating region for a full-scale installation requires: (1) the selection of a desired effluent quality; (2) determination of the θ_T/C^o value for this influent quality from the relationship of Figure 12; (3) selection of a value of C^o and the associated solids loading rate from the clarifier efficiency and loading relationships of Figure 14; and (4) calculation of the design θ_T value for the values of θ_T/C^o specified and C^o selected. The specification can then be completed using the above, the process kinetic constants and various conversion factors presented in Section VI, and material balance expressions presented in Section IV.

The selected treatment objective, i.e., the desired effluent quality in a full-scale application, will vary with factors such as effluent standards, additional treatment required, existing facilities available, etc. The treatment objective selected for exemplifying process specification subsequently is a design effluent soluble COD concentration of 29 gm/l, for which a θ_T/C^o value of 40 days is specified (Figure 12). This design "point" was selected because it represents the lower limit of the zone of transition in the operational relationship and because of the demonstrated stability of the anaerobic contact process at θ_T/C^o values in excess of 35 days. A C^o value of 0.375 (equivalent to a TSS removal efficiency of 62.5 percent), and a solids loading rate of 100 kg/day/sq m (approximately 20.5 lbs/day/sq ft) were selected as design values based on recommendations presented in Section VI. For the selected values of θ_T/C^o @ 40 days and C^o @ 0.375, the required hydraulic residence time, θ_T, is 15 days.

In the subsequent computations to complete the specification of the steady-state operations, the following relationships and parameter values were used (from data presented in Section VI):

1. Ratio of effluent soluble COD: total COD concentration = 0.95
2. Volatile fraction of MLTSS = 0.67
3. Methane production factor = 0.30 liter CH_4 per gm soluble COD removed
4. Average methane content of gas stream = 60 percent
5. Design influent COD concentration = 100 gm/l

The steady-state MLVSS concentration was computed using Equation 7, the values of the constant Y^o and K^D determined for the process, and the design values of θ_T and C^o as follows:

$$X_1^o = \frac{1{,}225 \ (100{,}000-29{,}000)}{[(0.129)\ (15)] + 0.375}$$

$$= 6{,}915 \text{ mg/l (MLVSS)}$$

For an assumed MLTSS volatile fraction of 0.67, the MLTSS concentration corresponding to the above MLVSS concentration is equal to 10,320 mg/l.

The rational organic loading rate for the design point and the above MLVSS concentration is:

$$= \frac{(100{,}000)}{(6{,}915)(15)}$$

$$= 0.964 \text{ kg soluble COD/day/kg MLVSS,}$$

and the corresponding volumetric loading rate is equal to 6.67 kg/day/cu m (0.416 lb/day/cu ft).

The methane gas production was computed as the product of the volumetric substrate reduction (gm soluble COD removed per liter of slops treated), and the methane production factor of 0.30 liter CH_4 per gm soluble COD removed, as follows:

Methane gas production:

$$= 0.30\ (100-29)$$

$$= 21.3 \text{ liter } CH_4 \text{ per liter slops treated}$$

The above production is equivalent to a methane production rate of 21.3 cu m CH_4/day per cu m/day of slops treated (2,850 cu ft CH_4/day per 1,000 gpd of slops treated). For an average methane content of 60 percent in the gas stream, the daily gas production rate is equal to 35.5 cu m per cu m/day of slops treated (4,750 cu ft/day per 1,000 gpd of slops treated).

The waste MLVSS production rate is equal to the difference between the rates of synthesis and autodestruction per unit volume of slops treated and is computed as follows:

Waste MLVSS production rate:

$$= Y^o \left[X_o^N - X_1^N\right] - K^D\theta_T X_1^o \qquad (17)$$

$$= 0.225(100{,}000-29{,}000) - 0.129(15)(6{,}915)$$

$$= 2{,}595 \text{ mg MLVSS per liter slops treated.}$$

For an assumed MLTSS volatile fraction of 0.67, then,

Waste MLTSS production rate

$$= 3{,}870 \text{ mg MLTSS per liter slops treated.}$$

The above is equivalent to a waste MLTSS production rate of 3.87 kg/day per cu m/day of slops flow treated.

The clarifier design requirements can be specified using the design C^o value, the associated solids loading rate, and the steady state MLTSS concentration. For a C^o value of 0.375, the design MLTSS removal efficiency is 62.5 percent; at a design MLTSS concentration of 10,320 mg/l, the estimated TSS concentration in the clarified effluent is 3,870 mg/l. For a design solids loading rate of 100 kg/day sq m, and the above MLTSS concentration of 10,320 mg/l, the corresponding liquid loading rate (overflow basis) is:

$$= \frac{100 \times 10^3}{10,320}$$

$$= 9.69 \text{ cu m/day/sq m (238 gpd/sq ft)}$$

OPERATING REQUIREMENTS

Three phases of operation need to be considered in the treatment of rum distillery slops by the anaerobic contact process: (1) startup; (2) routine operations; and (3) restartup after a seasonal shutdown of distillation operation. The factors of concern in each phase of operation are the operating requirements and the associated control and environmental variables available to the operator. Because the attainment and/or sustenance of steady-state conditions is a goal common to each phase of operation, the operating requirements can be developed in consideration of the preceding analyses of the steady state operating region.

Two basic premises inherent in the process kinetic theory, as found applicable in the present study for the characterization of rum distillery slops treatment by the anaerobic contact process, are that:

1. The effluent quality of the process is directly a function of the viable biomass in the system.
2. The only way that the biological process can respond to changes in influent quality, or more appropriately, to changes in the mass loading rate of biodegradable matter, is by an expansion or contraction of the viable biomass inventory in the system.

Corrollary to the preceding, the only way that process control can be effected is by control of the viable biomass inventory in the system.

From the perspective of operating requirements, the single transferrable parameter specifying the viable biomass inventory in the system is θ_T/C^o. In the present design example, the primary operating requirement is the maintenance of a θ_T/C^o value of 40 days at all times, if the design effluent quality is to be maintained.

From the analysis of the steady-state operating region presented above, the MLVSS concentration specified by a θ_T/C^o value of 40 days is equal to

6,900 mg/l. At the design hydraulic residence time, the required MLVSS inventory for an MLVSS concentration of 6,900 mg/l is equal to 103.7 kg MLVSS per cu m/day of slops treated (864 lbs MLVSS per 1,000 gpd of slops treated), and the attainment of this level of biomass inventory during startup, and its maintenance during routine operation and restartup, is the second operating requirement for the selected design condition.

The third operating requirement is the provision of temperature, pH, phosphorus, and mixing at levels specified below on the basis of operating experience (Section VI):

1. Temperature to be maintained at 35 ± 2°C
2. pH to be maintained at 7.2; the estimated lime dose required is 4,000 mg/l as CaO.
3. Phosphorus to be added to satisfy biological treatability requirements (the estimated dose required in 30 mg/l as P)
4. Complete and continuous mixing to be provided; the capacity of gas mixing provided in the pilot plant system (0.15 cu m/min/cu m of mixed liquor, or 20 SCFM/1,000 gallons) was sufficient to accomplish this.

Startup

The procedure recommended for startup on the basis of experimental operating experience is as follows:

1. Fill anaerobic fermenter with water; heat to operating temperature and maintain
2. Add seed sludge
3. Pretreat raw slops by neutralization to pH 7.2 and by phosphorus addition
4. Feed the pretreated raw slops at an initial rate not to exceed one kg soluble COD/kg MLVSS per day until the required biomass inventory (of approximately 100 kg MLVSS per cu m/day of total slops flow to be treated) is attained
5. Initiate polymer feed into the mixed liquor stream to the clarifier to maximize capture of MLVSS; waste no MLVSS during startup

The entire system should be monitored daily for gas production, alkalinity, pH, volatile acids, and temperature. Liquid, solids and nutrient balances should be performed at least weekly. The θ_T/C^o value should be calculated on a weekly basis to ensure that it equals or exceeds 40 days. The initial rate of feed of pretreated slops, as noted above should be adjusted upward or downward as the startup progresses, in consideration of system performance as tracked by the monitoring program.

The process of seed development took about four months, and the acclimation of the seeded units an additional four months in the present study. However, with the above guidelines for startup, developed from experience accrued in the course of this arduous effort, it is suggested that the entire process can be completed in half this period with the use of municipal digested primary sludge to supplement a slops-based seed sludge. Full-scale operational experience will be necessary before "standard practice" can be defined.

Routine Operation

Routine operation is defined as the maintenance of steady-state conditions within the framework of the above operating requirements. Routine chemical additions will be required as follows:

1. Lime, for pH control at a level of 7.2
2. Phosphorus, to ensure that phosphorus is not rate limiting
3. Polymer addition, to convert non-settleable solids to a settleable form

The entire system should be monitored as outlined above to establish the basis for:

1. Managing the solids inventory at a level commensurate with the mass organic loading rate
2. Controlling the dosage rates of the chemicals at the minimum necessary levels
3. Defining other necessary operating adjustments

Restartup

Restartup is defined herein as the time delay required to re-attain design load (steady-state) conditions after a period of interruption of slops flow due to seasonal production slowdowns or shutdowns. The factors associated with restartup are: (1) the duration of the shutdown or slowdown period; (2) the degree of reduction in slops flows during this period;

and (3) the time required to re-attain design loading after full production resumes.

The following options are available for dealing with the problem of minimizing the time delay.

1. Minimize the degree of reduction in feed supply by staggering the shutdown of sectors of production operation
2. Hold in reserve a quantity of feed to be used to restart the feeding of the system in advance of the resumption of full production
3. Use an alternative feed supply, such as raw municipal primary sludge, during the shutdown/slowdown period

Each of the above factors and options are situation-specific and will have to be examined on a case by case base to establish the economic tradeoffs.

The operating requirements presented above for startup and routine operation are equally applicable for restartup. These are:

1. The maintenance of the inventory of viable biomass in the system at the level specified by the selected design point (in the present case, approximately 104 kg MLVSS per cu m/day of total slops flow)
2. The control of the slops feed rate during restartup at a level not exceeding one kg/soluble COD/day/kg MLVSS
3. pH control at 7.2, and phosphorus addition
4. Temperature control at $35 \pm 2^{\circ}C$
5. Complete and continuous mixing, reinitiated upon the resumption of slops feed.

PROCESS FLOW SHEET AND DESIGN CRITERIA

The process flow sheet and design criteria for full-scale rum distillery slops treatment by the anaerobic contact process were developed in consideration of the preceding evaluations and the base of operating experience accured over the two-year period of investigation. The process flow sheet is presented in Figure 15, and the recommended design criteria are summarized in Table 10 and discussed below:

Unit Processes/Operations

The function of the slops storage tank is to provide equalizing of hour-to-hour flow variation, and to permit the cooling of the slops

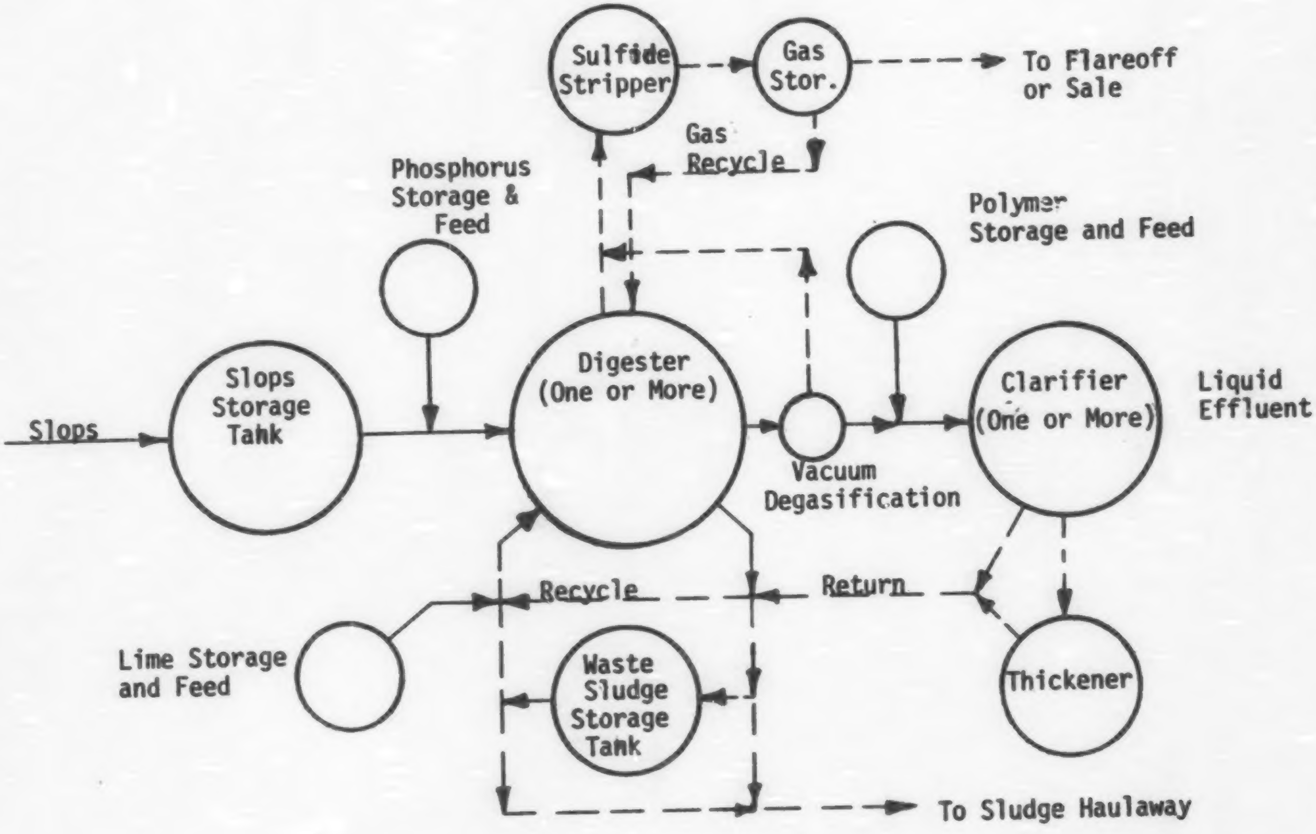

Slops
Slops Storage Tank
Phosphorus Storage & Feed
Digester (One or More)
Sulfide Stripper
Gas Stor.
To Flareoff or Sale
Gas Recycle
Polymer Storage and Feed
Vacuum Degasification
Clarifier (One or More)
Liquid Effluent
Recycle
Return
Lime Storage and Feed
Waste Sludge Storage Tank
Thickener
To Sludge Haulaway

FIGURE 15. PROCESS FLOW SHEET - RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS

TABLE 10. DESIGN CRITERIA FOR ANAEROBIC CONTACT PROCESS

Process Element	Recommended Design Criteria (Units as Noted)	
Slops storage tank	1 cu m per cu m/day slops flow	1,000 gal per 1,000 gpd slops flow
Digesters		
1. Capacity	15 cu m per cu m/day slops flow	2,000 cu ft per 1,000 gpd slops flow
2. Gas Mixing	0.15 cu m/min per cu m digester cap.	20 SCFM/1,000 gal digester volume
3. Heating	35 ± 2°C	95 ± 3°F
Clarifiers		
1. Liquid loading rate (overflow basis)	9.69 cu m/day/sq m	238 gpd/sq ft
2. Weir loading rate	12.4 cu m/day/m	1,000 gpd/ft
3. Recycle ratio	2:1	2:1
Thickeners		
1. Dry solids loading rate	122 kg/day/sq m	25 lbs/day/sq ft
Waste sludge storage tank	0.52 cu m per cu m/day slops flow	70 cu ft per 1,000 gpd slops flow
Gas handling system	36 cu m of gas per cu m/day slops flow	4,750 cu ft of gas per 1,000 gpd slops flow
Sulfide stripper	0.43 kg S per cu m/day slops flow	3.6 lb S per 1,000 gpd slops flow
Chemical feed systems		
1. Lime	4 kg CaO per cu m/day of slops flow	33.3 lbs CaO per 1,000 gpd slops flow
2. Phosphorus	0.3 kg P per cu m/day slops flow	2.5 lbs P per 1,000 gpd slops flow
3. Polymer	15 gm per cu m/day slops flow	0.125 lb per 1,000 gpd slops flow

stream emanating from the distillation columns to a temperature of 32 to 38°C. The excess sensible heat in this tank can be used to maintain the digester temperatures at 35 ± 2°C, as an option to using a portion of the methane gas production for this purpose. A minimum storage capacity of one cu m per cu m/day of slops flow (1,000 gal per 1,000 gpd slops flow) is suggested, exclusive of situation-specific storage requirements for restartup should the latter be required.

The digester capacity required at the selected hydraulic residence time of 15 days is 15 cu m per cu m/day of flow (2,000 cu ft per 1,000 gpd), and a minimum of two digesters should be provided for operational flexibility. The digesters should be heated to 35 ± 2°C. A gas mixing capacity of 0.15 cu m/min/cu m of digester volume (20 SCFM/1,000 gal) is suggested as a criterion for full-scale design based on the satisfactory pilot plant experience at this level.

The clarifiers should be sized to provide for a solids loading rate of 100 kg/day/sq m (20.5 lbs/day/sq ft) which, at an MLTSS concentration of 10,320 mg/l, is equivalent to a liquid loading rate (overflow basis) of 9.69 cu m/day/sq m (238 gpd/sq ft). A recycle ratio of at least 2:1 (ratio of slops feed flow: sludge recycle flow) is recommended based on the experience of Steffen and Bedker(14) in a slaughterhouse waste/anaerobic contact process application. For a recycle ratio of 2:1, the TSS concentration in the recycle stream is equal to 13,550 mg/l, and the design liquid loading rate (total basis) is equal to 29 cu m/day/sq m (713 gpd/sq ft). A design weir loading rate not to exceed 12.4 cu m/day/m (1,000 gpd/sq ft) is recommended on the basis of pilot plant experience.

In the specification of criteria for the thickener, it was assumed that: (1) the settled solids can be thickened at a loading rate of 122 kg/day/cu m (25 lbs dry solids/day/sq ft; and (2) the thickened sludge will have a dry solids concentration of 10 percent. These values were selected on the basis of design parameters typically associated with the thickening of primary sludges in municipal wastewater treatment plants.

The functions intended for the waste sludge storage are to: (1) provide waste sludge storage for a reserve sludge quantity equal to one-third of the biomass inventory in the digesters, to be available for transfer to the digesters as needed; and (2) to provide interim sludge storage prior to haulaway. The MLVSS biomass inventory required (103.7 kg MLVSS per cu m/day of slops heated), at a volatile fraction of 0.67, is equivalent to an MLTSS inventory of 155 kg MLTSS per cu m/day. For the storage of one-third of this inventory in the waste sludge storage tank, at an average concentration of 10 percent solids, the storage capacity required is equal to 0.52 cu m per cu m/day of slops treated (70 cu ft/1,000 gpd). At a waste MLTSS production rate 3.87 kg/day per cu m/day, the waste sludge storage capacity provided is sufficient to store 13 days of waste sludge production.

The gas handling system should be sized to handle 36 cu m of gas production per cu m/day of slops treated. Assuming that the gas stream contains 0.8 percent hydrogen sulfide by volume, then the sulfide stripper should be sized to handle 0.43 kg elemental sulfur per cu m/day of slops flow (3.6 lb per 1,000 gpd of slops flow).

Chemical feed systems are required for the addition of lime, phosphorus and polymer. Based on observations presented in Section VI, it is suggested that these systems be sized as follows:

1. Lime - feed at 4 kg as CaO per cu m/day of slops flow (33.3 lbs/1,000 gpd)
2. Phosphorus - feed at 0.3 kg as P per cu m/day of slops flow (2.5 lbs as P/1,000 gpd)
3. Polymer - feed at 15 gm polymer per cu m/day of slops flow (0.125 lb polymer per 1,000 gpd)

Layout

The layout as exemplified by the process flow sheet of Figure 15 incorporates the following material/liquid transfer capabilities:

1. From either digester to another or the same digester,
2. From the waste sludge storage tank to the digesters or to sludge haulaway,
3. From the digesters to the clarifier,
4. From the clarifier to the thickener, the waste sludge storage tank, and/or directly to the digesters.

The above transfer capabilities are required to support the types of material/liquid transfers required in the startup, routine operation, and restartup phases as discussed earlier in this section.

SECTION VIII

ECONOMIC ANALYSIS

The economic analysis presented in this section was developed using the following approach:

1. The design criteria of Table 10 and the process flow sheet of Figure 15 were used to develop estimates of capital and operating/maintenance costs at two design flow rates, 190 and 1,140 cu m/day (50,000 and 300,000 gpd);
2. The above cost estimates were developed using available unit process cost curves, updated to January 1974;
3. The heating requirements for the digesters at each design flow were estimated in consideration of ambient conditions in Puerto Rico, wherein most of the North American rum distilling capacity is located;
4. The value of the energy available as methane byproduct was assumed to be equal to the cost of purchasing an equivalent amount of energy in the form of fuel oil; and,
5. The cost estimates at each design flow rate were converted to unit treatment costs ($/volume treated) from which a cost capacity relationship was then constructed.

CAPITAL COSTS

The estimated capital costs for anaerobic contact treatment systems at design capacities of 190 cu m/day (50,000 gpd) and 1,140 cu m/day (300,000 gpd) are presented in Table 11. These cost estimates were developed from cost data presented by Eilers and Smith(22) and Patterson *et al*(23), updated to January 1974 cost levels, and by vendor contact. Overhead (engineering and administration) and contingency costs were estimated at 38 percent of the construction cost.

The estimated total construction costs were $1,683,000 for a 190 cu m/day facility, and $5,596,000 for a 1,140 cu m/day facility. The corresponding unit total construction costs are: $8,860 per cu m/day of capacity ($33,660/1,000 gpd) at the design flow rate of 190 cu m/day;

TABLE 11. CAPITAL COSTS (JANUARY 1974)

Item	Capacity @ 190 cu m/day (50,000 gpd)	Capacity @ 1,140 cu m/day (300,000 gpd)
Slops storage tank[1]	32,000	112,300
Mechanical mixers - slops storage tank	12,000	42,000
Control & solids handling building[2]	175,000	634,000
Lime storage, slaking, and feeding system	21,000	78,000
Phosphorus storage, dissolution and feeding system	3,000	10,000
Polymer storage and feeding system	1,500	5,000
Sulfide stripper	70,000	200,000
Gas storage/handling	90,000	270,000
Anaerobic digesters[3]	400,000	1,400,000
Vac. degasification	50,000	170,000
Gravity separator	65,000	125,000
Gravity thickener	38,000	80,000
Waste sludge storage	8,600	28,000
Pumps	12,500	46,000
General work site work, yard piping, general elec. & HVAC	240,000	855,000
Construction Cost	1,219,100	4,055,300
Overhead; contingency	463,200	1,541,000
Total construction cost	1,683,200	5,596,300

[1] Concrete in place @ $275/cu yd

[2] Includes areas for control room, lime storage, slaking, feeding, sulfide stripper and gas handling equipment, phosphorus storage, dissolution and feeding equipment, polymer storage, dissolution and feed equipment, pumps, etc.

[3] Concrete @ 275/cu yd; covers; heat exchanger.

and $4,910 per cu m/day ($18,650/1,000 gpd) at the design flow rate of 1,140 cu m/day.

OPERATING AND MAINTENANCE COSTS

The estimated annual O/M (operating and maintenance) costs for anaerobic contact systems at design capacities of 190 cu m/day and 1,140 cu m/day are presented in Table 12. Labor costs were estimated at a rate of $5.40 per man hour, including overhead. The chemical costs were estimated as follows: lime at $17.50/ton; phosphorus at $50/ton; and polymer at $1/lb. Sludge haulaway costs were estimated at a unit cost of $0.30/ton dry solids/mile of haul, assuming a 25 mile haul. Electricity costs were estimated by assuming a pumping requirement of 0.004 HP per 1,000 gpd of slops flow(22), and an annual cost of $330/yr/HP, the latter based on pump efficiencies of 60 percent and electricity costs of $0.03/KWH.

Fuel costs for digester heating are not included in Table 12 on the assumption that a portion of the methane production would be allocated for this purpose. An evaluation of the allocation required, and of the net methane production after the allocation is made, is presented below.

Exclusive of fuel cost for heating, the estimated annual O/M costs are $39,500 at the design capacity of 190 cu m/day (50,000 gpd), and $150,400 at the design capacity of 1,140 cu m/day (300,000 gpd).

ECONOMIC VALUE OF METHANE PRODUCTION

The factors considered in determining the economic value of the methane production were:

1. The gross methane production,
2. The allocation required, from the gross methane production, to heat digesters;
3. The net methane production, i.e. the excess available after deduction of the digester heating requirement from the gross methane production.

The first step in the evaluation was the estimation of digester heating requirements at each scale. The requirements were estimated using the procedure of Babbitt and Baumann(24) and the following

TABLE 12. ANNUAL OPERATING AND MAINTENANCE COSTS (JANUARY 1974)

Item	Capacity @ 190 cu m/day (50,000 gpd)	Capacity @ 1,140 cu m/day (300,000 gpd)
Labor	16,200	54,000
Materials and supplies (except chemicals)	8,000	20,000
Chemicals	6,000	35,400
Electricity	6,800	27,000
Other utilities	500	2,000
Sludge haulaway	2,000	12,000
Total O/M Cost[a]	39,500	150,400

Note: [a]excludes heating cost.

assumptions (assuming location of the facilities in Puerto Rico):

1. Average ambient temperature: 21°C (70°F)
2. Digester feed stream and sludge recycle stream temperatures: 27°C (80°F)

On this basis, the estimated BTU requirements for digester heating were: 2.45×10^9 BTU/year at the design capacity of 190 cu m/day; and 13.8×10^9 BTU/year at the design capacity of 1,140 cu m/day. Given that a barrel of fuel oil has a BTU equivalent of approximately 6,000,000 BTU, the preceding BTU requirements for digester heating are equivalent to fuel oil consumption rates of: 408 bbl/yr at a design capacity of 190 cu m/day; and 2,300 bbl/yr at a design capacity of 1,140 cu m/day.

The fuel oil equivalent of the gross annual methane production was estimated using the following:

1. The gross methane production factor of 21.3 cu m CH_4 per cu m slops treated (2,850 cu ft per 1,000 gallons treated) as presented in Section VI

2. An energy equivalent of 1,000 BTU/cu ft of methane, or 5.90 bbl of fuel oil/1,000 cu m of methane (0.167 bbl of fuel oil/1,000 cu ft of methane)
3. An assumed annual production schedule of 300 days

For these conditions, the fuel oil equivalent of the gross annual methane production is 37.7 bbl/year per cu m/day of capacity (143 bbl/year per 1,000 gpd); i.e., is equal to 7,160 bbl/year at a design capacity of 190 cu m/day (50,000 gpd), and 42,900 bbl/year at design capacity of 1,140 cu m/day (300,000 gpd).

The allocation of gross methane production required for digester heating purposes can be evaluated by comparing the ratio of fuel oil equivalent consumption for digester heating with the above fuel oil equivalent production rates, at each design scale. This ratio is equal to 408/7,160 (5.7 percent) at 190 cu m/day capacity, and 2,300/42,900 (5.4 percent) at 1,140 cu m/day capacity. Thus, at either scale, less than six percent of the gross methane production must be allocated for digester heating in the Puerto Rico environment, and the net methane production is equal to at least 94 percent of the gross methane production. At 94 percent, the fuel oil equivalent of the net methane is equal to 35.4 bbl/year per 1,000 gpd). On a dollar basis, this fuel equivalent is worth $0.119 per cu m of slops treated (0.45/1,000 gallons treated) for each $/bbl of fuel oil cost on the open market.

TREATMENT COSTS

Annual treatment costs ($/cu m or $/1,000 gal of slops treated) were estimated on an unadjusted basis (assuming a fuel oil value of $0/bbl) and on an adjusted basis assuming fuel oil values up to $28/bbl. A summary of the unadjusted annual costs at each design flow rate is presented in Table 13. The amortization costs were developed assuming an interest rate of eight percent, a 15-year life for 40 percent of the capital investment, and a 25-year life for 60 percent of the capital investment.

The unadjusted total annual cost of the full-scale applications is $212,800 at the design capacity of 190 cu m/day (50,000 gpd), and $726,500 at 1,140 cu m/day (300,000 gpd). For a 300-day production schedule, the

TABLE 13. UNADJUSTED ANNUAL COSTS FOR RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS

Item	Capacity @ 190 cu m/day (50,000 gpd)	Capacity @ 1,140 cu m/day (300,000 gpd)
Capital costs[1]	$1,683,200	$5,596,300
Annual costs ($/yr)		
. Amortization[2]	173,269	576,083
. O/M costs[3]	39,500	150,400
. Total annual cost[4]	212,769	726,483
Unit treatment costs[5]		
. Per cu m treated	3.74	2.13
. Per 1,000 gallons treated	14.18	8.07

[1] From Table 11

[2] Amortization at 8% interest rate; 15 year life for 40% of capital investment and 25 year life for 60% of capital investment

[3] From Table 12

[4] Exclusive of methane credit

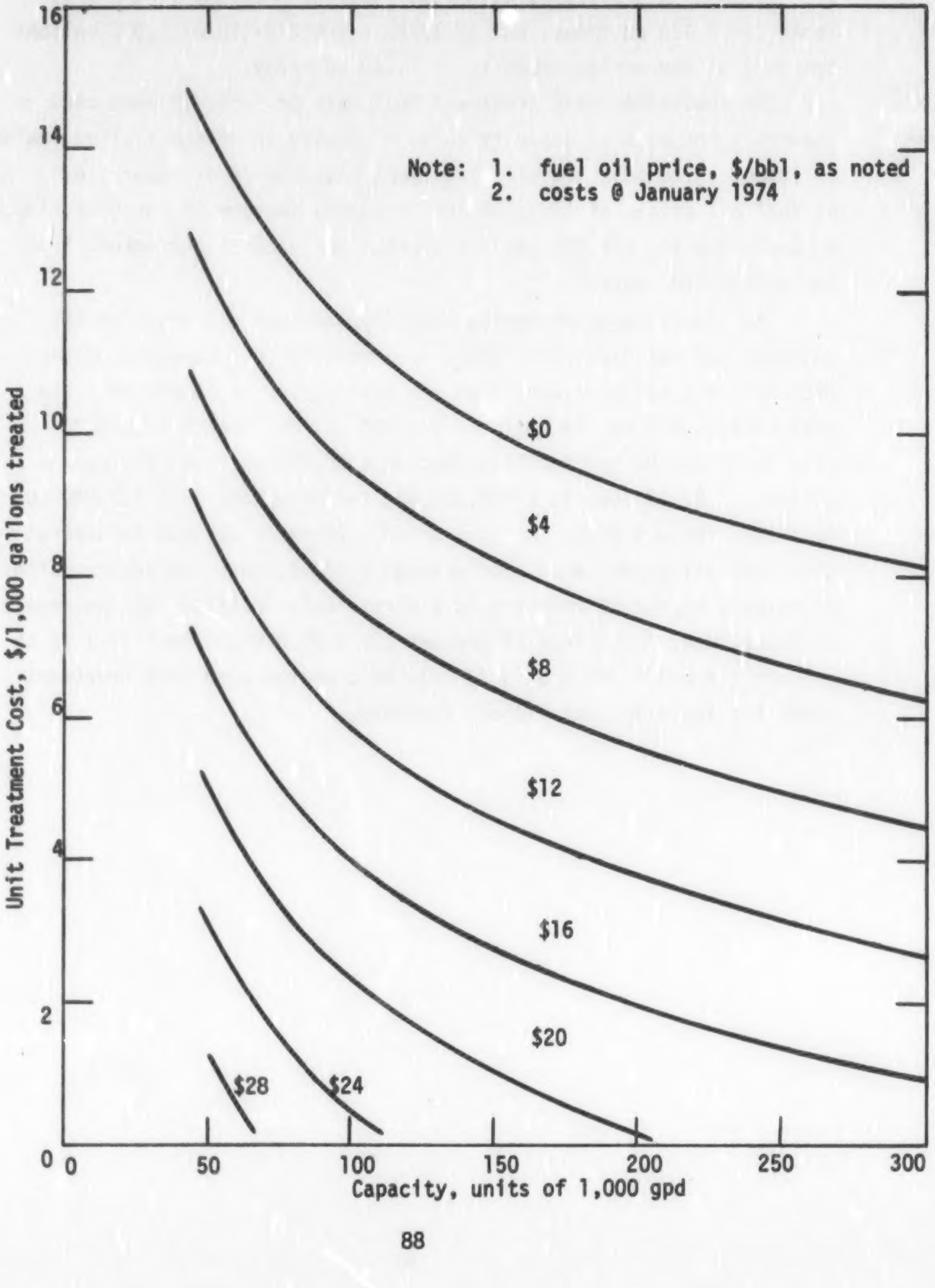
[5] Based on 300 day production schedule

unadjusted total annual costs are equivalent to unit treatment costs of: $3.74/cu m treated ($14.18/1,000 gallons treated) at the design capacity of 190 cu m/day; and $2.13/cu m treated ($8.07/1,000 gallons treated) at the design capacity of 1,140 cu m/day.

The unadjusted unit treatment cost data of Table 13 were used to construct the cost vs capacity curve of Figure 16 at the fuel oil value of $0/bbl. The adjusted unit treatment cost curves of Figure 16, at fuel oil prices of $4 to 28/bbl in $4/bbl increments, were developed by deducting $1.80/1,000 gallons treated for each $4 increment, from the unadjusted costs.

The significance of design capacity and fuel oil price on the unit cost of rum distillery slops treatment by the anaerobic contact process are clearly evident from the cost curves of Figure 16. The unit treatment cost (in units of $/1,000 gallons treated) is $.80 less in a 100,000 gpd facility than in a 50,000 gpd facility and an additional $3.30 less in a 300,000 gpd facility than in a 100,000 gpd facility, for any given fuel oil price. If it is assumed (conservatively) that fuel oil prices will soon average $12/bbl, then the incorporation of methane byproduct recovery in a plant-scale installation can reduce unit treatment costs from 35 percent (in a 50,000 gpd facility) to 65 percent (in a 300,000 gpd facility), as compared with unit treatment costs for installations without recovery.

FIGURE 16. UNIT TREATMENT COSTS FOR RUM DISTILLERY SLOPS TREATMENT BY ANAEROBIC CONTACT PROCESS



SECTION IX

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SECTION X

LIST OF PUBLICATIONS

The following publication is pending:

1. Shea, T. G., and G. H. Dorion, "Investigation of Rum Distillery Slops Treatment by Anaerobic Contact Process," presented at Fifth National Symposium on Food Processing Wastes, April 17-19, 1974.

SECTION XI

GLOSSARY

A listing of the nomenclature used in this report is presented in Table 14. In addition to this nomenclature, the following abbreviations have been used:

1. BOD - five day, 20°C, biochemical oxygen demand.
2. B/S - bench-scale
3. CH_4 - methane
4. COD - chemical oxygen demand
5. MLTSS-mixed liquor total suspended solids
6. MLVSS-mixed liquor volatile suspended solids
7. P/P - pilot plant
8. TSS - total suspended solids
9. VSS - volatile suspended solids

TABLE 14. NOMENCLATURE

K = a constant equivalent to the substrate concentration when $k = k^m/2$

K^D = specific autodestruction rate, time^{-1}

k = specific growth rate, time^{-1}

Y^o = mass of organisms produced/mass of substrate removed, mg/mg

X^o = concentration of organisms, mg/l

X_o^o, X_1^o, X_2^o = influent; reactor effluent, and separator effluent organism concentrations, mg/l, respectively

X_w^o = mass of organisms leaving system per unit time/total flow leaving system per unit time, mg/l

$X_R{}^o$ = recycle stream (and sludge wastage stream) organism concentration, mg/l

C^o = recycle factor = $X_w^o/X_1{}^o$

X^N = concentration of controlling substrate, mg/l

X_o^N, X_1^N, X_2^N = influent, reactor effluent, and separator effluent substrate concentrations, mg/l, respectively

$X_R{}^N$ = recycle stream substrate concentration, mg/l

C'' = degree of concentration of substrate in the solids separator ($=X_R{}^N/X_1{}^N$)

θ = time

θ_T = theoretical hydraulic detention time, time

Q = influent flow rate, l/day

Q_s = sample flow rate, l/day (average)

Q_w = sludge wastage flow rate, l/day

r = fraction of influent recycled

SECTION XII

APPENDIX

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APPENDIX A

DATA PROCESSING PROCEDURES

The two steps required in the processing of the information base developed for each biological system, as derived from the monitoring effort described in Table 2, were as follows:

1. The flow-weighted averaging of all daily parameter values obtained on a weekly base, in the determination of average weekly flow rates, and average weekly VSS (volatile suspended solids) and soluble COD concentrations entering, within, and leaving the anaerobic fermenter and the solids separator of each system.
2. The development, with the weekly averaged data base, of flow, substrate (soluble COD), and solids balances in the format necessary for development of the parameters from which the plots of Figures 2a and 2b could be developed.

The discussion below deals with the mechanics of the second step, i.e., the development of the required parameters.

The development of the paramters was done as follows:

1. The weekly averaged data base for each system was manipulated to develop the values of the parameters, $\left[X_o^N - X_1^N \left[1-r(C''-1)\right]\right] / C^o X_1^o$ and θ_T / C^o, necessary for determining the values of the kinetic constants and K^D and Y^o by the graphical procedure of Figure 2a.
2. The steady-state criteria described in Section VI were then applied to the weekly averaged data base, and the values of the above two parameters, to identify the weeks of operation with each system deemed representative of steady-state.
3. The resultant values of the above parameters as associated with steady-state operation were then plotted for the determination of the values of Y^o and K^D (as presented in Figure 10 of Section VI).
4. The value of K^D as established in the preceding step was then used in the computation, for the steady-state data, of the values of the parameters, $\theta_T/(K^D\theta_T + C^o)$ and $1/X_1^N$, necessary for determining the values of the kinetic constants k^m and K by the graphical procedure of Figure 2b.

5. The values of the above parameters were then plotted for the determination of the values of k^m and K, as is presented in Figure 11 of Section VI.

The computational procedures used in determination of the weekly values of the parameters, $\left[X_o^N - X_1^N \quad [1-r(C''-1)]\right]_o /C^oX_1^o$ and θ_T/C^o involves the computation first of the value of C^o using Equation 12, and then of the values of the parameters using the calculated value of C^o. This procedure is exemplified below, using weekly averaged data for the 22nd week of pilot plant operations.

(1) Weekly averaged data base

Q = 82.1 liters/day (slops feed rate)

Q_w = 0 (no sludge wastage)

Q_s = 0

r = 6.08

θ_T = 20.6 days

X_1^o = 5,200 mg/l (VSS)

X_2^o = 1,700 mg/l (VSS)

X_o^N = 67,050 mg/l (soluble COD)

X_1^N = 22,770 mg/l (soluble COD)

X_R^N = 21,190 mg/l (soluble COD)

(2) Determination of C^o using Equation 12:

$$C^o = \frac{0 + (82.1 - 0 - 0)(1,700)}{(82.1)(5,200)} + \frac{0}{82.1}$$

$$= 0.327$$

(3) Determination of $\left[X_o^N - X_1^N \quad [1-r(C''-1)]\right] /C^oX_1^o$:

$$= \frac{67{,}050 - 22{,}770\left[1-6.08\left(\frac{21{,}190}{22{,}770} - 1\right)\right]}{(0.327)(5{,}200)}$$

$$= 20.38 \text{ mg soluble COD/mg VSS}$$

(4) Determination of θ_T/C^o:

$$= \frac{20.6}{0.377}$$

$$= 63.0 \text{ days}$$

A summary of the data used in the computation of the parameters $\theta_T/(K^D\theta_T + C^o)$ and $1/X_1^N$ is presented in Table 15. The weekly values of the former parameters were determined using $K^D = 0.0667$ days^{-1} (from Figure 10). The parameter values reported in Table 15 were used to develop the plot of Figure 11 for the determination of the kinetic constants k^m and K.

TABLE 15. SUMMARY OF PARAMETERS FOR DETERMINATION OF k^m AND K - STEADY STATE DATA

System	Week	θ_T (days)	C^o	X_1^N (mg/l)	$\frac{\theta_T}{K^D\theta_T+C^o}$ (days)	$\frac{1}{X_1^N} \times 10^5$ $(mg/l)^{-1}$
P.P.	21	25.2	0.411	18,950	12.05	5.28
	22	20.6	0.327	22,770	12.10	4.39
	24	19.0	0.359	21,569	11.69	4.64
	25	20.3	0.525	23,813	10.80	4.20
	26	19.3	0.558	24,422	10.46	4.09
	33	31.0	0.417	26,257	12.48	3.81
	34	36.9	0.492	29,452	12.50	3.40
1	25	35.1	0.604	21.230	11.92	4.71
	26	35.4	0.523	21.250	12.28	4.71
2	29	221	1.004	17,785	14.03	5.63
	30	149	0.992	16,456	13.63	6.07
	31	147	1.000	16,790	13.61	5.96
	32	143	1.012	15,109	13.55	6.62
	33	145	1.002	14,504	13.59	6.89

Note: $K^D = 0.0667$ days^{-1} (from Section VI)